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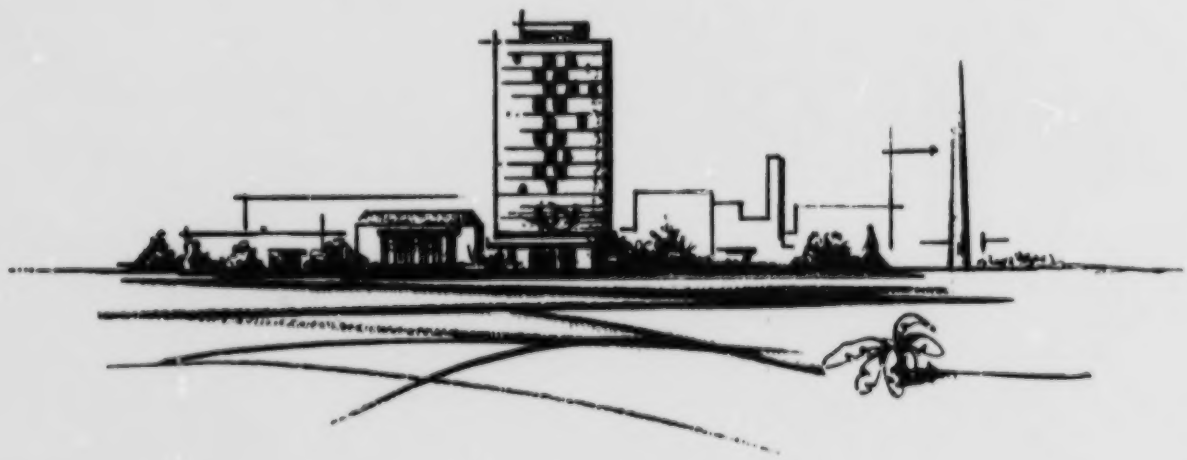
RESEARCH REPORT

FEASIBILITY STUDY OF CENTRALIZED
AIR-POLLUTION ABATEMENT

to

NATIONAL AIR POLLUTION
CONTROL ADMINISTRATION
PUBLIC HEALTH SERVICE
DEPARTMENT OF HEALTH, EDUCATION,
AND WELFARE

November 17, 1969



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FINAL REPORT

on

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CONTROL ADMINISTRATION
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by

M. Fels and H. L. Crawford

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INTRODUCTION

Industrial plant complexes discharge a variety of pollutants to the environment. These discharges include both particulate and gaseous pollutants over a wide range of concentration, volumetric flow rates, and temperatures. Recently, there has been increasing concern regarding the esthetic and health aspects of these emissions to the atmosphere, and it has become necessary to limit the amount of pollutants discharged to the atmosphere. Unfortunately, the devices to control pollution are expensive, and, in many cases, the pollutants collected are of no economic value to the company. Thus, any reduction in the economic burden of air-pollution control would be greatly welcomed by both government and industry alike.

The concept investigated in this report stems from an attempt to develop less expensive means of air pollution control. This concept is the centralization of pollution control by having the polluted effluents from individual plants come to a centralized treatment facility. The reasoning behind this concept is that one large piece of control equipment would cost less per unit amount of gas treated than would several small units. For example, in the chemical-process industry, capital costs of equipment increase, in general, by size ratio to the 0.6 power, thus favoring larger units.

If implemented, the centralized concept would provide a system whereby all polluted gaseous effluents could be discharged, untreated, to a system of manifolds which would permit the gases to be pumped to the centralized treatment facility. At the centralized facility, the total mixed polluted gases would be purified and discharged to the atmosphere.

The purpose of the present study is to present the results of a preliminary feasibility study of the centralized air-pollution-abatement concept.

SUMMARY

The technical and economic aspects of a centralized air-pollution-control plant located at a distance from seven industrial plants were investigated. The plants chosen were: (1) lime, 200 tons/day, (2) cement, 4500 barrels/day, (3) sulfuric acid, 400 tons/day, (4) power, 25 Mw, (5) fertilizer, 570 tons/day, (6) gray iron, 1440 tons/day, and (7) electric arc, 2600 tons/day. Gaseous- and particulate-emission levels were taken from literature sources, and as far as possible, average values were used for each industry.

The total amount of gas from these plants was found to be 627,000 cfm at 320 F, after mixing. The study showed the following results:

	Capital Cost, million dollars	Operating Cost, millions of dollars/ year
Individual Treatment	5.4	1.7
Centralized Abatement		
Pollution Control	2.5	0.9
Transportation (3 miles)	11.5	2.8

It was concluded that, although the centralized control facility is less expensive than individual control devices, the transportation cost is so high as to make the centralized concept unattractive. The results also show that the economics would begin favoring the centralized concept if each of the seven plants were located at about 1/2 mile from the central facility; this distance was considered to be unrealistically close from the standpoint of individual plant's land requirements.

Other disadvantages of the centralized concept included: (1) emissions from the lime, cement, and sulfuric plants were rendered valueless, (2) a malfunction in the centralized control equipment would create large problems by releasing large quantities of pollutants over a relatively small area, (3) vegetation growth over the buried gas pipes would be inhibited, leading to potential esthetic problems.

APPROACH

As originally conceived, the study was to utilize air-pollution-emission data in a document entitled "The Kanawha Valley Report" to develop criteria for (1) selection of a model site for the feasibility study and (2) performance of the economic and technical analyses necessary to provide a basis for judging the concept feasibility.

The Kanawha Valley Report was received in draft form and reviewed carefully. It was found that the type of data obtained in the Kanawha Valley study and the method of reporting these data rendered the results essentially useless for the research program. Specifically, the individual company questionnaire requested data on plant emissions in terms of pounds per day; but, for estimating clean-up costs, gaseous flow, and pollutant concentrations are required. In addition, data from the individual company surveys are not reported. Instead, total estimated emissions are reported as tons of pollutants discharged per year in the entire study area.

Therefore, it was decided to attempt to collect the necessary data from individual companies. These data would take the form of composition of effluent gases, flow rates, temperatures, and the means and economics of pollution control in present use. In the course of discussions with industry representatives, a number of difficulties associated with this approach were uncovered, which led to its abandonment.

A decision was then made to work with a hypothetical industrial complex using emission data which could be found in the literature. A brief preliminary study showed that there were enough emission data for the purposes of this study from the following industries:

(1) Cement Plant
(2) Lime Plant
(3) Sulfuric Acid Plant
(4) Coal-Burning Power Plant
(5) Fertilizer Plant
(6) Gray-Iron Foundry
(7) Electric-Furnace Steel Plant

The discussion to follow deals with the emission characteristics and pollution-control costs for each of the seven plants. Consideration of the total abatement concept will be done by discussion of the system for transporting the effluent gases to the centralized treatment plant, and of the central plant itself.

CHARACTERISTICS OF INDUSTRIAL PLANTS CHOSEN

Cement Plant

Process Description

Portland cement is made by mixing calcareous (calcium-containing) and argillaceous (clay-containing) materials in the proper ratios. Essentially, the unit operations prepare the raw materials in the necessary proportions and in the proper physical state of fineness and intimate contact so that the chemical reactions can take place at the temperature in the kiln to form various silicates and aluminates. Four major steps in the production of portland cement are quarrying and crushing, grinding and blending, production, and finished grinding and packaging. A good description of these operations is given in Kreichelt, et al.[1]

Cement is made by either the wet process or the dry process. In the dry process, the raw materials are ground, mixed, and blended as dry powders; whereas, the wet process involves a slurrying of the raw materials in the grinding, mixing, and blending operation.

Emissions and Their Control

Emission Data. Particulate matter is the primary emission in the manufact e of portland cement. There are also the normal combustion products of the fuel used to supply heat for the kiln and drying operations, including oxides of nitrogen and small amounts of oxides of sulfur. The three sources of dust emission are the crushing operation, grinding operation (in the dry process), and the kiln operation.

The purpose of the crushing operation is to reduce the quarried material from boulder-size down to convenient size, 3/4-inch rock for handling and transportation to the cement plant. This operation is usually done at the quarry. The amount of dust generated during this operation depends on factors such as moisture content of the rock, methods of handling, and type of rock. Because of the remote location of the operation, and because of the relatively large sizes of rock involved, the dust emissions are of minor importance. Thus, for this study, these emissions were neglected.

The purpose of the grinding operation is to reduce the size of the cement rock to a fineness suitable for effective reaction in the kiln. This operation is performed on the dry material in the "dry process", and a considerable amount of dust is generated. In the wet process, water is added to the grinding mill with the crushed feed to form a slurry. No dust is generated in the grinding operation because of the slurry condition. Kreichelt, et al., note that there are 110 wet-process plants and 69 dry-process plants.[1] Therefore, emission calculations were made for the wet process, and dust arising from the grinding operation was not taken into account.

The dust from the kiln is of major importance. Generated by the drying and clinkering of the cement itself, this dust is entrained by the gaseous products in the combustion of fuel at one end of the kiln. Kreichelt, et al., give emission data for 22 wet-process cement plants; from these data, averages were calculated for use in the present study. Calculations for the gaseous composition are shown in the Appendix A. Table 1 shows the data used in subsequent calculations for emission control costs.

TABLE 1. EMISSION DATA FOR CEMENT PLANT KILN

4500 Barrels/Day

Variable	Value
Gas Volume, 1000 acfm	193
Gas Temperature, F	404
Particulate Loading, gr/ft^3	5.4
Outlet Loading, gr/ft^3	0.05
Gas Composition, volume percent	
N_2	48.2
O_2	1.3
CO_2	19.5
H_2O	31
SO_2	Trace

After clinkering, the cement is ground again to a fineness of about 10-micron average particle size. Closed circuit grinding is the common practice; dust collectors that are an integral part of the fine-grinding equipment keep atmospheric emissions down to a negligible level.

Emission Control. The emissions from the cement plant are in the form of relatively fine particles; for example, from 20 to 40 percent of the dust has a diameter less than 5 microns. Consequently, Kreichelt, et al.[1], report that only collectors such as the electrostatic precipitator effectively collect fine dust. According to the data given for several plants, 8 of 14 dry-process plants use electrostatic precipitation, and only 1 wet-process plant of 22 does not use electrostatic precipitation for pollution control. Therefore, in this study it was assumed that the cement plant will use electrostatic precipitation as its primary means of pollutant dust control.

Lime Plant

Process Description

Lime is produced by heating limestone to decompose the limestone into carbon dioxide and lime. Heat is produced by burning coal or natural gas. Both vertical and rotary kilns are used in the production of lime, but rotary kilns have a higher capacity. A good discussion of the lime industry is given by Lewis and Crocker.[2]

Emissions and Their Control

Emission Data. The major air contaminant from lime manufacturing is dust. This dust arises from crushing, screening, and kiln discharge. The dust consists mainly of $CaCO_3$ and CaO, which can be irritating to the eyes, respiratory membranes, and moist skin. Although some dust is produced from the crushing and screening operations, the major emissions are generated in the kiln operation.

Typical dust loadings for both the vertical- and rotary-kiln operation are given by Lewis and Crocker.[2] Also given are typical data for production rates, gas temperatures and volumes, fuel-to-lime ratios, and CO_2 content of the effluent gases. On the basis of these data, calculations for the composition of the exit gases can be made; they are shown in Appendix A. Because rotary kilns are more common and also create more of an air-pollution problem, the rotary lime kiln was chosen as the basis for a typical plant in the centralized air-pollution-abatement model. Table 2 gives the pertinent data necessary on emissions from the lime plant.

TABLE 2. EMISSION DATA FOR LIME PLANT

200 Tons/Day

Variable	Value
Gas Volume, 1000 acfm	29
Gas Temperature, F	910
Particulate Loading gr/ft^3	6
Gas Composition, volume percent	
N_2	56.5
O_2	1.4
CO_2	31.8
H_2O	10.2
SO_2	0.1

Emission Control. In general, four types of control equipment are used to clean kiln gases. They are cyclone collectors, bag filters, electrostatic precipitators, and water scrubbers. From data given in Lewis and Crocker, it appears that the bag filters do a satisfactory job of particulate removal from effluent gases.[2] Also, they are relatively inexpensive compared with other control equipment such as the electrostatic precipitators and water scrubbers. Cyclone collectors can be used only as primary elements for the removal of the plus 10-micron-size dust. Thus, for this preliminary study, costs were based on the use of bag filters in the lime plant. Other data on pollution control in the lime industry can be found in References (3) and (4).

Sulfuric Acid Plant

Process Description

Sulfuric acid is made generally by combustion of sulfur with air to yield SO_2; oxidizing the SO_2 catalytically to SO_3; and reaction of SO_3 with water to yield the final product, H_2SO_4. Two processes by which H_2SO_4 is produced are the chamber and the contact processes.

In the chamber process, SO_2 is oxidized to SO_3 in the presence of various oxides of nitrogen. The oxidation takes place in large chambers, and the nitrogen oxides are then recovered in a "Guy Lussac" tower. The chamber process currently produces approximately 10 percent of the total sulfuric acid in the United States, but it is expected to account for less in the future. This process yields relatively weak acid. Because the acid is more dilute than acid from the contact process, transportation costs per unit of H_2SO_4 are higher. Construction and operating costs are usually higher than those for contact plants. For these reasons the chamber process is today a small factor in sulfuric acid production.

In the contact process the gases leaving the combustion chamber (8 to 11 percent SO_2) are cooled and passed over a solid catalyst which promotes the oxidation of SO_2 to SO_3. The SO_3 is then absorbed in water in an absorption tower. Cuffe and Dean give a detailed description of the processing operation involved.[5]

Emissions and Their Control

Emission Data. The major source of air pollution in the sulfuric acid industry is the discharge of SO_2 and acid mist in the effluent gas stream. SO_2 in the stack gas results from the incomplete conversion of SO_2 to SO_3 in the catalyst converter. Conversion efficiencies of 98.0 to 98.5 percent are attainable with proper plant design. Higher conversion efficiencies require a more expensive plant and result in higher production costs. The unconverted SO_2 from the catalyst converter passes through the absorption system and is discharged to the atmosphere. Data of Cuffe and Dean show that SO_2 concentrations in the absorber discharge stack ranged from 0.13 to 0.54 percent, with a mean (based on 33 tests) of 0.26 percent.[5]

Sulfuric acid mists in stack gases arise from two sources. The first is related to the hydrocarbons present in sulfur (from 0.1 to 0.3 percent). When sulfur is burned, the hydrocarbons present are converted to CO_2 and water. The water later reacts with the SO_3 produced in the converter to form sulfuric acid. The reaction takes place in the vapor phase with the resulting sulfuric acid condensing out as a fine aerosol or mist as the gas cools down in equipment and/or ducts between the converter and absorbing tower. Much of the mist formed in this manner passes through the absorber packing and must be collected by mist-collection equipment, if objectionable mists are to be eliminated.

The second source is sulfuric acid spray which generally results from the entrainment of acid by the gas on leaving the packing in the drying and absorbing towers.

The pertinent data for pollution control calculations were obtained from Cuffe and Dean[5], and are shown in Table 3. Calculations for the amount of nitrogen and oxygen content of the gas are shown in Appendix A.

TABLE 3. EMISSION DATA FOR SULFURIC ACID PLANT

400 Tons/Day

Variable	Value
Gas Volume, 1000 acfm	26.5
Gas Temperature, F	154
Acid Mist Loading, gr/ft^3	0.13
Gas Composition, volume percent	
N_2	85.8
O_2	13.8
SO_2	0.26

Emission Control. Although the SO_2 amounts to 0.26 percent in the stack gases, Cuffe and Dean report that little or no recovery is attempted in actual practice.[5] Presumably, one of the reasons for this is the relatively low flow (about 26,000 cfm) of effluent gases from an average sulfuric acid plant. Thus, the SO_2 discharge, in terms of amount per hour, is quite low. However, in a few plants, scrubbers have been used to reduce the SO_2 level in the stack gases. As there are no reliable cost data for SO_2 removal for sulfuric acid plants, it will be assumed for the purposes of this study that SO_2 control will not be used. The exchange of the mist eliminator (discussed below) for a wet scrubber would not change the economics substantially.

Much more work has gone into the investigation of acid-mist removal. Brink has developed a fiber mist eliminator which can operate at efficiencies of about 99 percent.[6, 7] In this design, the gas containing the mist is passed through a fiber packing.

The mist particles are collected as a film on the surface of the fibers. The collected liquid flows through the bed by gravity, drips into a seal pot, and is recycled to the process. In this way, essentially all of the sulfuric acid mist can be collected.

Power Plant

Process Description

Basically, the power-plant concept as applied here is concerned with the combustion of a fuel to produce steam; the steam then drives turbogenerators for the generation of electricity. The majority of plants utilize coal as a fuel, but oil or gas are also burned to some extent.

Emissions and Their Control

Emission Data. The major pollutants from fossil-fuel-burning plants are oxides of sulfur and nitrogen. Sulfur dioxide, the major constituent, is the prime object of current and future control activities.

The sulfur oxides in the combustion gases arise from combustion of the sulfur in the fuel. For example, the sulfur content of coal in this country ranges from about 0.5 to about 6 percent by weight.

The gaseous emissions from a typical power-generating station are from 50 to 100 times that of the total of the other 6 processes considered in this study. Therefore, for this study a small plant (25,000 Kw) is used; one that would, for example, supply the small industrial complex used in this study. Gas-flow rates and compositions were obtained from Katell[8] and are shown in Table 4.

TABLE 4. EMISSION DATA FOR POWER PLANT

25,000 Kw

Variable	Value
Gas Volume, 1000 acfm	70
Gas Temperature, F	300
Particulate Loading, gr/ft^3	1.5
Gas Composition, volume percent	
N_2	76.2
O_2	3.4
CO_2	14.2
H_2O	6.0
SO_2	0.2
SO_3	Trace

Emission Control. Many processes have been proposed to deal with the removal of sulfur dioxide from stack gases. They are all in various stages of development from bench scale to large pilot-plant operations. Two processes, the catalytic-oxidation and the dolomite process, have been developed enough for reasonable cost estimation[9], and they show some promise of commercial feasibility.

In the catalytic-oxidation process, the hot flue gases are cleaned of their particulate matter by electrostatic precipitation, and then are passed into equipment which is essentially a contact sulfuric acid plant. In the dolomite process, the flue gases are contacted with a mixture of carbonates, mainly calcium and magnesium (either dry or in slurry form). Reaction of SO_2 with the dolomite occurs, producing the corresponding sulfates, which are then disposed of.

The catalytic oxidation process requires a greater investment than does the dolomite process, but it does not have the disposal problems of the dolomite process. Economic aspects of both of these processes are discussed in the section on "Economics of Individual Air Pollution Control". Because of the amount of effort in the area of SO_2 control devices, better systems probably will be evolved; as shown later, such improvements would merely strengthen the conclusions of this study.

Fertilizer Plant

Process Description

Fertilizers, as defined here, are those containing phosphorus. Common to all these fertilizers is the starting material, phosphate rock. Phosphate rock is mined in open pits, then screened and washed to recover the larger size fraction. Subsequent flotation recovers the smaller particles of additional value. The wet rock is then dried and ground for shipment and further processing. The major fraction is shipped as ground phosphate rock.

The fertilizer products to be considered here are run-of-pile triple superphosphate, diammonium phosphate, and granular triple superphosphate. Run-of-pile triple superphosphate is produced by the continuous acidulation of dried and ground phosphate rock with phosphoric acid in a reactor, such as a TVA mixing cone. The resulting slurry is discharged directly to a slow moving settling belt where the reaction produces a solid material. This "green" triple is stored from 30 to 60 days in a curing building where the reaction continues at a decreasing rate. When cured, the product will contain about 46 percent soluble P_2O_5. The final product is mined from the "pile" in the curing shed. It is crushed, screened, and shipped in bulk.

Some 5 to 6 years ago, a new fertilizer product known as "diammonium phosphate" was introduced to the market. It is produced by the ammoniation of phosphoric acid with a subsequent granulation and drying. The acid is preneutralized in a vertical reaction tank with additional ammoniation and granulation taking place in rotating horizontal drums. The product is dried in rotary-drum kilns, then sized and transferred to the storage and shipping building.

Two different methods are used for the manufacture of granular triple superphosphate. One method continuously acidulates phosphate rock with phosphoric acid. The

resulting slurry is fed to blungers where it is mixed with recycle materials to start the granulation. From the blungers the material is passed through granulating drier kilns at a controlled speed and temperature. The other method passes cured and screened run-of-pile material through drum granulators in the presence of steam. It is then dried in horizontal, rotary kilns.

Emissions and Their Control

Engdahl and Sachsel describe the sources of pollution arising from fertilizer-plant operation.(10) Two sources of pollutant exist in fertilizer manufacture: rock dust from grinding, and fluoride-bearing fumes and gases from acidulation.

If the air-swept mill and the dust collector for rock grinding are properly designed, over grinding is the most likely source of dust that can usually be traced to improper operation of the mill.

Acidulation fumes and gases cannot be avoided. In the dried and ground state, phosphate rock contains 3.5 to 4.0 percent fluoride, of which 1/3 to 1/2 is evolved from subsequent chemical or thermal processing. The evolved fluorides are probably present as silicon tetrafluoride and hydrogen fluoride.

Huffstutler reports emission levels for the various fertilizer plants in pounds per hour of fluorides and ammonia in the case of diammonium phosphate.(11) These data are summarized as averages in Table 5. This source does not give quantitative data on total exhaust gas rate. It was assumed that the exhaust from the buildings would be approximately 100,000 cfm at ambient temperatures.

TABLE 5. EMISSION DATA FOR FERTILIZER PLANT

Fertilizer Product	Average Production, ton/day	Emissions Before Scrubbing, lb/ton	Emissions After Scrubbing, lb/ton	Scrubber Efficiency, percent
Run-of-Pile Triple Superphosphate	860	5.8	0.16	97.2
Diammonium Phosphate	644	6.2	0.38	92.7
Granular Triple Superphosphate	214	3.8	0.15	96.0
Averages	573	5.3	0.23	95.3

Emission Control. As mentioned before, the major source of pollution is in fluoride emissions. Fortunately, SiF_4 and HF are quite soluble in water. Therefore, a wet-contact device can be used to remove these fluorides. Huffstetler reports the use of wet scrubbers in the fertilizer industry. Recent information suggests that the scrubber design called a "crossed flow packed tower" is more suitable for fluoride

removal than a venturi scrubber. However, as there are no cost or efficiency data on this type of device, it is assumed that a wet scrubber of the venturi type will be adequate for pollution control of fluorides.

Gray-Iron Foundry

Process Description

The principal device used in iron foundries to obtain molten metal for production of castings is the cupola furnace. A cupola can be defined simply as a refractory-lined cavity with necessary openings at the top for the escape of gases and for the charging of the stock, and openings at the bottom for entry of the air blast and for drawing off the iron and slag.

Cupolas are used for almost all gray-iron melting, and as the primary melting unit for much malleable-iron production. They are used commercially in heats lasting from 2 hours a day to several successive days, with hourly outputs from 2 to 50 tons. Additional information on cupola operation can be obtained from Reference (13).

Emissions and Their Control

Emission Data. Emissions from the cupola consist of coke and flux particles, various metals, their oxides, and some condensible oils and greases. Emissions of sulfur compounds are usually small because the sulfur content of cokes is generally 0.6 percent or less.

Because of the variations of infiltrated air, grain loadings also show wide variations. For example, data on three cupolas published in Reference (13) show variations in grain loadings from 0.0014 to 1.6 gr/ft^3. For the present study, the highest grain loading was taken, and grain loadings assumed to be about 2 gr/ft^3.

Calculations for the gaseous emission compositions are shown in Appendix A, and the data are tabulated in Table 6. Cupolas are operated as either "hot-blast" or "cold-blast". In the hot-blast operation, the air for combustion of the coke is preheated by the exiting stack gases. The system also incorporates an afterburner to burn off the CO and H_2 emissions; the heat of combustion of these two emissions is used to help heat the blast. The afterburner probably also oxidizes the small amount (<0.001 gr/ft^3) of condensible oils and greases that are released during the cupola operation.

Most large, modern cupolas are hot-blast operations; however, a relatively large number of small cold-blast units are still in operation. In these units, ambient air is used for the blast, and the stack gases are discharged without heat recovery.

If a cold-blast operation were in use, an afterburner would be necessary for the removal of the CO and oils, whether the emissions are treated at the cupola site or are treated in the centralized facility. However, for the centralized concept, a cooler would be necessary to cool the stack gases before they are transmitted to the centralized

facility. The cost of the cooler would have to be added to the cost of the centralized treatment plant. It does not seem feasible to discharge the cupola emissions from a cold-blast operation directly into the centralized facility. The condensibles would probably make frequent cleaning of equipment necessary, and the removal of the relatively large amount of CO from the mixed gases would be difficult and expensive. Therefore, for this study, a hot-blast operation was assumed to avoid undue penalizing of the central facility concept.

TABLE 6. EMISSION DATA FOR GRAY-IRON FOUNDRY

60 Tons/Hour

Variable	Value
Gas Volume, 1000 acfm	49.6
Gas Temperature, F	500
Particulate Loading, gr/ft^3	2
Gas Composition, volume percent	
N_2	78.2
CO_2	19.7
O_2	0.4
H_2O	1.7
SO_2	Trace

Emission Control. Bloomfield has noted that collection equipment being purchased for cupolas is of five types: (1) wet caps, (2) fabric filters, (3) low-pressure-drop wet scrubbers, (4) high-pressure-drop wet scrubbers, and (5) multitube cyclones.(14)

Wet emission-control systems are generally used for cupolas because the high temperatures of cupola effluents restrict uses of other methods.

Wet-cap systems operate with natural draft. Pressure drop is about 1 inch of water. They are made of stainless steel with rubber liners and can become part of a wet-scrubber system in a future improvement. They remove only large, heavy particles.

The use of fabric filters requires cooling of cupola gases before contact with the filter. Electrostatic precipitators used with cupolas have failed to maintain high collection efficiencies because of wide variations of gas-feed conditions, i.e., temperature, humidity, volume and particulate chemical composition. Life and dependability of these precipitators has been low and no attempts have been made in recent years to employ them.

Sterling states that installation of wet scrubbers can accomplish a very high degree of control.(15) For this study, therefore, it was assumed that a wet-scrubber installation would be chosen.

Electric-Arc Furnace

Process Description

The electric-arc furnace consists of a refractory-lined shell to hold the material to be melted, electrodes that can move vertically, and the means to tilt the pot. The electric arc is characterized by its high temperature and concentration of heat energy. The electrodes are either graphite or amorphous carbon. The heat is provided by an arc to the charge or melt. Because the source of heat is nonchemical, electric furnaces are especially desirable in melting alloys of controlled composition.

In 1967, electric arc furnaces produced about 11 percent of the total raw carbon steel made in the United States and 36 percent of the alloy and stainless steels. About 59 percent of all electric furnace heats were carbon steels. Electric-arc furnaces have capacities up to 200 tons.

Emissions and Their Control

Emission Data. Emissions from the electric-arc furnace originate from light scrap that oxidizes readily, from dirty scrap (a major source), and from oxygen lancing. The main emissions are fumes from scrap preheating, iron oxide dust from the melting operations, and furnace off-gases.

The amount of dust released per net ton of electric-furnace steel depends upon the condition of the scrap and whether or not oxygen lancing is used. Dirty scrap can raise the dust emissions from a normal level of 8 to 15 pounds to as high as 40 pounds per net ton of steel. It has been estimated that oxygen lancing produces 20 percent of the total emissions. The composition of the off-gas from the electric furnace varies with practice. The chief constituents are carbon monoxide, carbon dioxide, nitrogen, and oxygen. A more detailed description of emissions from the electric-arc furnace can be found in Barnes and Lownie.[16] Table 7 lists the emissions from the electric-arc furnace, calculated from data in Barnes and Lownie (Appendix A).

TABLE 7. EMISSION DATA FOR ELECTRIC-ARC FURNACE

2600 Tons/Day

Variable	Value
Gas Volume, 1000 acfm	116.0
Gas Temperature, F	500
Particulate Loading, gr/ft^3	2.5
Gas Composition, volume percent	
N_2	78.5
O_2	13.5
CO_2	1.35
CO	6.75

Emission Control. The types of air-pollution-control devices that can be used in the electric-arc furnace are: high-energy venturi scrubbers, electrostatic precipitators, and fabric filters. Of these, the fabric filter is most often used, probably because it is relatively inexpensive compared to the first two. Therefore, for this study, a fabric-filter installation was chosen to control the air-pollution problems from the electric-arc furnace.

ECONOMICS OF INDIVIDUAL AIR-POLLUTION CONTROL

In the ensuing discussions of the costs of air-pollution-control equipment, it must be emphasized that accurate estimates both of the capital and operating costs involved are virtually impossible. There are several basic reasons for this:

(1) Each installation of a control device is different and depends on many factors, such as actual plant geometry, labor, etc. Thus, cost information from seemingly similar installations in many cases shows discrepancies of up to 50 to 100 percent.

(2) The cost of operating the control installation is dependent on a great many factors. Even the cost of the device in the plant itself varies from time to time. Some of these factors include age of installation, changes in production rate or in feed materials, etc.

(3) It is often very difficult to obtain cost information from the industries using these devices.

The following section details the cost estimates for each of the seven plants. When there were several sources of cost data for a particular type of installation, the cost figures were averaged. Also where a range was quoted, the mid-point of the range was taken.

One general assumption which was made was that the annual cost could be estimated by taking the operating costs and adding 20 percent of the capital expenditures.[16] Again, it must be kept in mind that these costs can be in error by 50 to 100 percent. However, as will be seen later, errors of this magnitude do not affect the final conclusion of this study. Costs quoted here have been scaled to 1969 prices. Table 8, which appears at the end of this section, summarizes the data on the costs.

Cement Plant

As mentioned previously, the pollution-control device of choice is the electrostatic precipitator. The cost of an electrostatic precipitator is a function of many parameters, the major ones being the gas load and efficiency. The efficiency is in turn related to the particle size distribution. Sargent gives data on electrostatic precipitator costs for a 60,000 cfm (68 F) unit.[17] He also lists the efficiencies of this precipitator with a "standard dust". With his data it is possible to calculate that the efficiency of the electrostatic precipitator whose costs are given would be about 92 percent when

handling the dust emissions from the cement plant. As the efficiency required is (5.4-0.05)/5.4 = 99 percent, the electrostatic-precipitator costs reported by Sargent would be too low; and, if this precipitator were used, the outlet loading would be about 0.4 gr/ft^3. Data taken from Barnes and Lownie (page C-80) indicate that reducing the dust loading 2-1/2 times requires a 10 percent increase in capital and operating expenditures. From this information, it can be calculated that the capital and operating costs reported by Sargent should be increased by about 24 percent. Pertinent calculations are detailed in Appendix B.

The cost data of Sargent yield a capital expenditure of $490,000 and a yearly operating cost of $106,000. Barnes and Lownie report a capital cost of $905,000 and an annual operating expense of $277,000 for an electric-arc furnace handling 185,000 cfm at 500 F. It was assumed that a reasonable estimate of the costs involved could be obtained by averaging the costs reported by both Sargent and by Barnes and Lownie.

Lime Plant

Lewis and Crocker report a capital expense of $1.80 per cfm and annual operating expenses of $.20 per cfm.[2] Bergstrom et al., report these costs to be $1.55 and $.08 (maintenance only).[4]

Data of Sargent[17] indicate capital and annual operating expenses (including 20 percent of capital) of $1.50 and $.47 per cfm (at 910 F). Barnes and Lownie[16] give costs of $2.13 and $.78 per cfm for capital and annual operating expenses for a fabric filter to control particulate matter from electric furnaces.

When the 20 percent of capital expense is added to the operating costs reported by Lewis and Crocker and by Bergstrom, an average of all the costs yields $1.75 per cubic foot capital and $.55 per cubic foot annual operating expense.

Sulfuric Acid Plant

To control mists, a fiber mist eliminator will be used. In order to obtain the cost of these units, Dr. Brink[18] was contacted, and he provided the cost curves for the mist eliminators (see Figure B-1 in Appendix B). A conservative estimate was made by choosing the most costly unit. The price of this unit for the gas-flow rate of 26,500 cfm, is about $55,000. Brink suggests that the installation costs would be about 100 percent of the unit costs. Direct operating expenses would be low and, thus, the operating cost per year was assumed to be 20 percent of the capital cost.

Power Plant

As mentioned before, the two processes to be considered for SO_2 control are the catalytic oxidation and the dolomite process. Capital and operating costs for the catalytic oxidation process have been detailed by Lemmon et al., based on the work of

Katell.[8] For a 25 Mw unit, capital investments would be $1.9 million, with annual operating expenses of $630,000. This operating expense assumes the saleability of sulfuric acid at $14 per ton. However, in the present case, sulfuric acid production would amount to only about 15 tons/day, and the price obtainable per ton at this small production level would probably be lower, resulting in a higher operating expense.

For the dolomite process, the slurry, or wet process was chosen because it has a higher efficiency and can remove 90 percent of the SO_2. Schuler, et al., have given data on both capital and annual operating costs of the wet dolomite process.[19] These costs for the 25 Mw power plant are $645,000 and $250,000, respectively. Thus, it can be seen, that, for the relatively small power plant used in the present study, the dolomite process would be the preferred means of SO_2 control.

Fertilizer Plant

Costs for a high-energy, venturi-type scrubber having an efficiency of over 99 percent as reported by Sargent[17] are $180,000 capital, and $90,000 annual operating. Barnes and Lownie[16] report costs for a high-energy, wet scrubber which handles 100,000 cfm at 100 F to be $357,000 and $168,000 for capital and annual operating, respectively. For the fertilizer plant, a low-energy scrubber installation would probably be adequate. Thus, cost data reported by Barnes and Lownie were lowered by the ratio of the cost of high-energy wet scrubbing to low-energy wet scrubbing as computed from the data of Sargent.[17] Again, the average of the two sets of data was taken to give an estimate of the costs involved. Calculations are shown in Appendix B.

Gray-Iron Foundry

For the gray-iron foundry, costs of venturi scrubbers were estimated from the data of Barnes and Lownie.[16] The gaseous emissions from the electric-arc furnace, open-hearth furnace, and basic oxygen furnace have similar characteristics as those from the cupola. Barnes and Lownie report capital costs for wet scrubbers on the three furnaces ranging from $7.20 to $9.30 per cfm, and annual operating expenses ranging from $3.00 to $3.72 per cfm. Average values were $8.03 and $3.39 per cfm, respectively.

Electric-Arc Furnace

Costs for the electric-arc furnace control are well documented by Barnes and Lownie.[16] From their data for fabric filters (pages C-43 and V-20), capital costs were estimated to be $343,900, and operating costs to be $195,000 per year.

TABLE 8. SUMMARY OF INDIVIDUAL AIR-POLLUTION-CONTROL COSTS

Industry	Gas Flow Rate, acfm	Temp, F	Pollution-Control Device	Equipment Capital Investment, dollars	Equipment Operating Expense, dollars/year
Cement Plant	193,000	404	Electrostatic Precipitator	745,000	206,000
Lime Plant	29,000	910	Fabric Filter	50,800	16,000
Sulfuric Acid	26,500	154	Mist Eliminator	55,000	11,000
Power Plant	70,000	300	Wet Limestone	645,000	250,000
Fertilizer Plant	100,000	90	Wet Scrubber	210,000	99,800
Gray-Iron Foundry	49,600	500	Wet Scrubber	400,000	168,000
Electric-Arc Furnace	116,000	500	Fabric Filter	343,000	195,000
Totals	583,000			2,449,000	946,000

Stack, Plant, and Labor

Stack Costs

Cost data for the stacks, which includes the column itself, the supports, and linings were taken from data given by Stankiewicz.[20] Installation was estimated to be 100 percent of the equipment cost; and as before, maintenance, depreciation, etc., expenses were taken to be 20 percent of the capital cost on a yearly basis.

As the discharges from the stacks should be relatively clean, it was felt that 200-foot stacks should suffice. Table 9 summarizes the data on stack costs.

TABLE 9. STACK-COST DATA

Plant	Stack Diameter, ft	Stack Velocity, ft/sec	Capital Cost, dollars	Operating Expense, dollars/year
Cement	10	41	121,800	24,400
Lime	5	25	77,800	14,600
Sulfuric acid	5	23	77,800	14,600
Power	5	60	77,800	14,600
Fertilizer	7.5	38	100,000	20,000
Gray iron	5	43	77,800	14,600
Electric arc	7.5	44	100,000	20,000

Plant Costs

Not included in the cost calculations, so far, are costs for plant, buildings, services, etc. Lang has published average values by which the capital-cost estimate should be increased to include buildings, piping, services, etc.[21] For solids- and fluid-processing plants under $1 million (Table 8), an additional 75 percent is to be added to the capital cost. These costs are detailed in the Summary Table (Table 14, page 29).

Labor

Labor involved to run the individual pollution-control devices was estimated to require about 1/2 man, and to run the centralized facility, 2 men. Hourly rates were assumed to be $4.50, which includes management charges. The labor costs appear in the Summary Table (page 29).

At this point, the major capital and operating costs for pollution control of the seven plants have been detailed. The following two sections are concerned with the costs of the gas-transport system and the plant necessary to remove the pollutants from the total mixed effluent gases from the seven plants.

GAS TRANSPORTATION SYSTEM TO CENTRAL FACILITY

The detailed cost discussion that follows is divided into three sections – piping costs, fan costs, and motor costs – to facilitate later use of the information in studies of many different pipe/fan/prime mover combinations. A summary of assumptions used appears at the end of this section.

In the study, it was assumed that all of the seven different types of industrial plants considered were located at various equal distances from the central processing plant. Costs involved for different distances over a range from 1 to 5 miles were estimated. It was assumed that the central plant would eliminate polluting dirt and chemicals, before expelling the collected air into the atmosphere.

A single pipe line from each industrial plant to the processing plant with no branches was analyzed, for convenience; however, pipe networks might be used, if the pollutants transported were compatible when mixed. Figure 1 is a schematic diagram of the transportation system.

Three arrangements of fan locations relative to the processing plant were considered for analysis, but only one was selected. Two fan pressure levels were considered, the lower pressure level being for the shortest pipes.

The three arrangements of fan locations were as follows:

(1) One fan located in each pipe at each industrial plant for blowing polluted air through the pipe to the processing plant

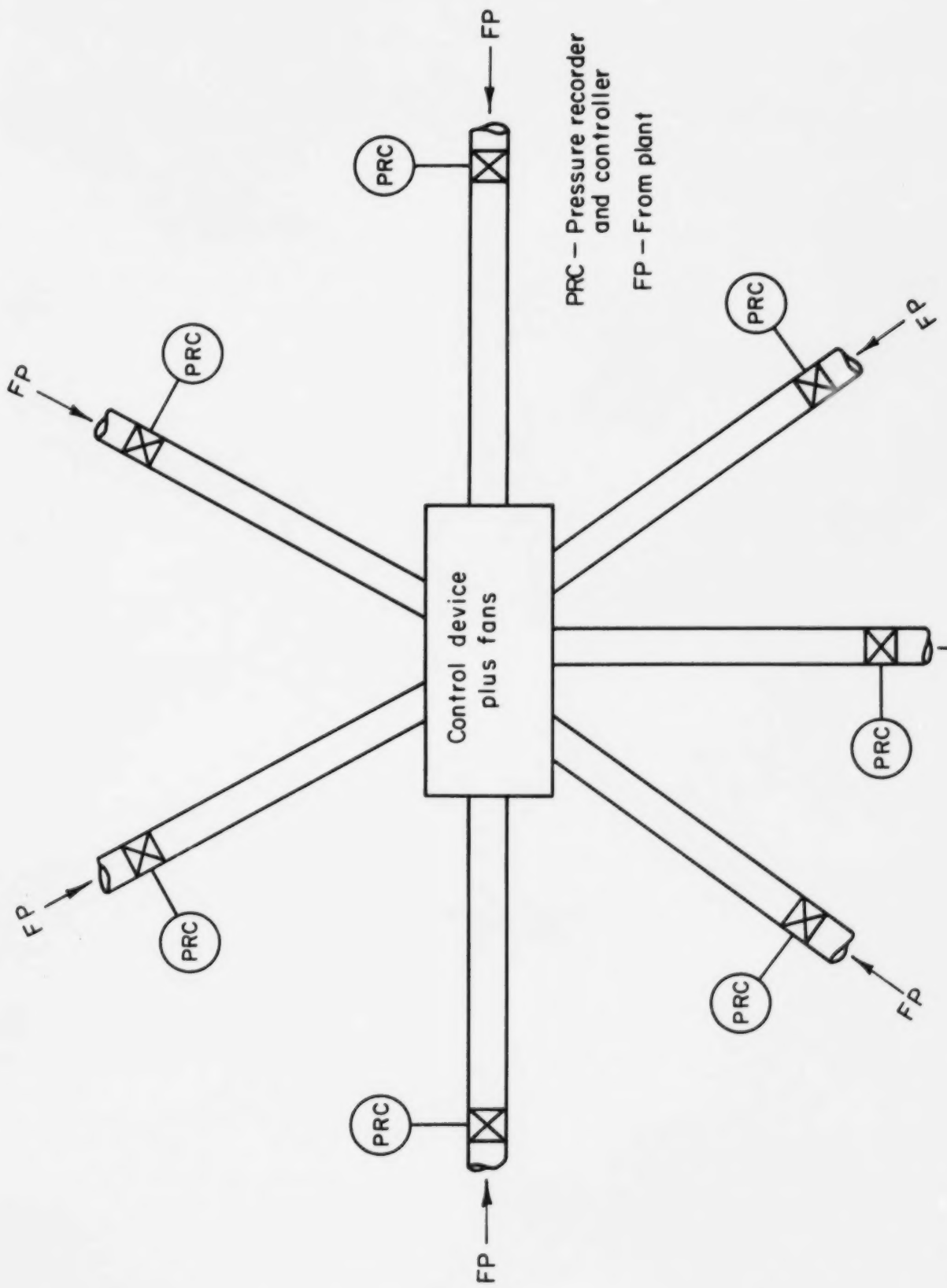

FIGURE 1. PLOT PLAN OF GAS TRANSPORT SYSTEM

(2) One fan located in each pipe just ahead of the processing plant for inducing polluted air through the pipe and for blowing polluted air through the processing plant

(3) Several large fans located after the processing plant for inducing polluted air through the entire system.

The last arrangement listed is recommended and is the only one analyzed herein. The major advantage of placing the fans downstream of the processing plant is the reduction of fan costs realized by handling clean, low-temperature air. Erosion and corrosion are reduced substantially by removing dirt particles and chemicals from the polluted air before passing it through the fans; therefore, carbon steel rather than expensive alloys can be used for fan construction. The handling of low-temperature air also eliminates the need for high-temperature-resistant alloys. The capacity requirement of each fan is reduced significantly at low temperature, since the air density is relatively high. This reduces the size and, hence, the initial cost of the fans required as well as the cost of power for operating the fans.

Another advantage of this arrangement is that the pressure in the piping is below atmospheric; therefore, any small holes that might develop in the pipes would leak air into them. If such leaks occurred in the positive-pressure system of the first arrangement listed above, pollution of the atmosphere would result from the outflowing gases.

Minimum transport velocities were calculated for the particle loading of each industrial plant. The minimum transport velocity, which occurs when particles no longer slide along the bottom of the pipe but are transported by saltation, has been recommended by Thomas, et al.[22], as the optimum design velocity for suspension transport. The particle size, particle specific gravity, and the minimum transport velocity of the particle for each industrial plant are shown in Table 10. The air velocities of only two of the six plants having particle-laden air are sufficiently high to suspend the particles. Since the air velocity was limited to 100 feet per second by erosion considerations, particles of the sizes and weights assumed for the electric-steel furnace, gray-iron foundry, lime plant, and cement plant will settle to the bottom of the pipe. Whether these particles will slide along the bottom of the pipe or collect to block the air flow is not known, but the problem does not appear to be insurmountable.

TABLE 10. MINIMUM TRANSPORT VELOCITY FOR POLLUTED AIR

Industrial Plant	Maximum Expected Particle Diameter, microns	Particle Specific Gravity(a) (Water = 1.0)	Minimum Transport Velocity, ft/sec
Electric Power	150	1.6	79.1
Electric Steel Furnace	100	7.72	166.1
Gray-Iron Foundry	100	7.12	251.8
Lime	150	2.5	155.2
Sulfuric Acid	50	1.0	69.5
Cement	60	2.5	130.4
Fertilizer	No Particle Loading Assumed		

(a) Mechanical Engineer's Handbook, edited by T. Baumister, Sixth Edition, McGraw-Hill, New York (1958).

A periodic cleaning with a rotating plug, such as that used in the pipe-laying industry, might be a workable solution to the problem. Another alternative might be the installation of simple, low-efficiency dust collectors at each industrial plant. These would be relatively inexpensive and would operate on less than 2 inches of water-gage pressure loss.

Piping Cost Estimates

Cost estimates were made for piping for a range of distances from 1 to 5 miles between the centralized processing plant and the circle containing the industrial plants.(23 through 28) Piping costs for a given length are determined approximately by the diameter and material. Figure 2 shows the variation of cost for several diameters of carbon steel pipe. Costs vary linearly with distance and diameter. For example, the cost of a 2-mile-long pipe is twice that of a 1-mile-long pipe and the cost of an 8-inch-diameter pipe is twice that of a 4-inch-diameter pipe of the same length and material. Therefore, the cost of a 4-inch-diameter pipe 1 mile long was chosen as the base value for computing all piping costs. The basic value, $30,000 per mile for carbon steel pipe, for example, includes the cost of material, installation, and easement rights for land. Material costs include the cost of seamless pipe, coating material around the pipe (no thermal insulation), cathodic protection, and one damper in each line for flow control. Installation costs include surveying, mapping, field welds of pipe, freight, sales taxes, field-labor fringe benefits and statutory burdens, field supervision, temporary facilities, equipment, tools, and contractor fees. Additional details can be found in Appendix B.

It was assumed that the pipe would be buried underground. Methods of installing the piping other than burying underground were considered, but a detailed economic study was not carried out. These methods were to install the piping above streets or buildings, install it in cradles just above ground, and laying it on the ground partially exposed. These methods were rejected on the basis of thermal pollution, safety hazards of hot exposed piping to personnel, appearance reasons, and vulnerability to sabotage by pranksters. The degree of undesirability would depend considerably on location of the installation, local codes, and community standard, habits, and acceptance.

Pipe costs for three materials are shown in Figure 3, for a given pipe size. Carbon steel (ASTM-A-53, Grade A) is recommended where temperatures are below 800 F and where excessive corrosion is not a problem. Chrome-molybdenum steel (ASTM A-335, Grade P12) is recommended at 800 F to 1000 F(29) which exceeds the highest expected temperature of this study. Stainless steel (ASTM A-312, Type 316) is recommended where special protection against corrosion is needed.(29, 30, 31, 32) Other pipe materials such as plastics and concrete, were considered, but, generally, temperatures were above the range of application of these materials.

The most economical pipe size is generally the smallest in diameter; however, as the diameter is decreased, the air velocity in the pipe approaches certain limits. Two limits on velocity were used herein. The upper velocity limit of 100 ft/sec(30) was selected so that erosion, caused by dirt particles suspended in the air, will not be "excessive".* A second limit is that the frictional pressure loss in the pipe must not

*Reference 30, Table 4, pp 18-8, shows the "Maximum Allowable Gas Velocity...to prevent flue dust erosion" for pulverized coal of 100 ft/sec and cement dust of 45 ft/sec. This illustrates that more information concerning erosion is needed before a final decision is made on the maximum velocity in an actual installation.

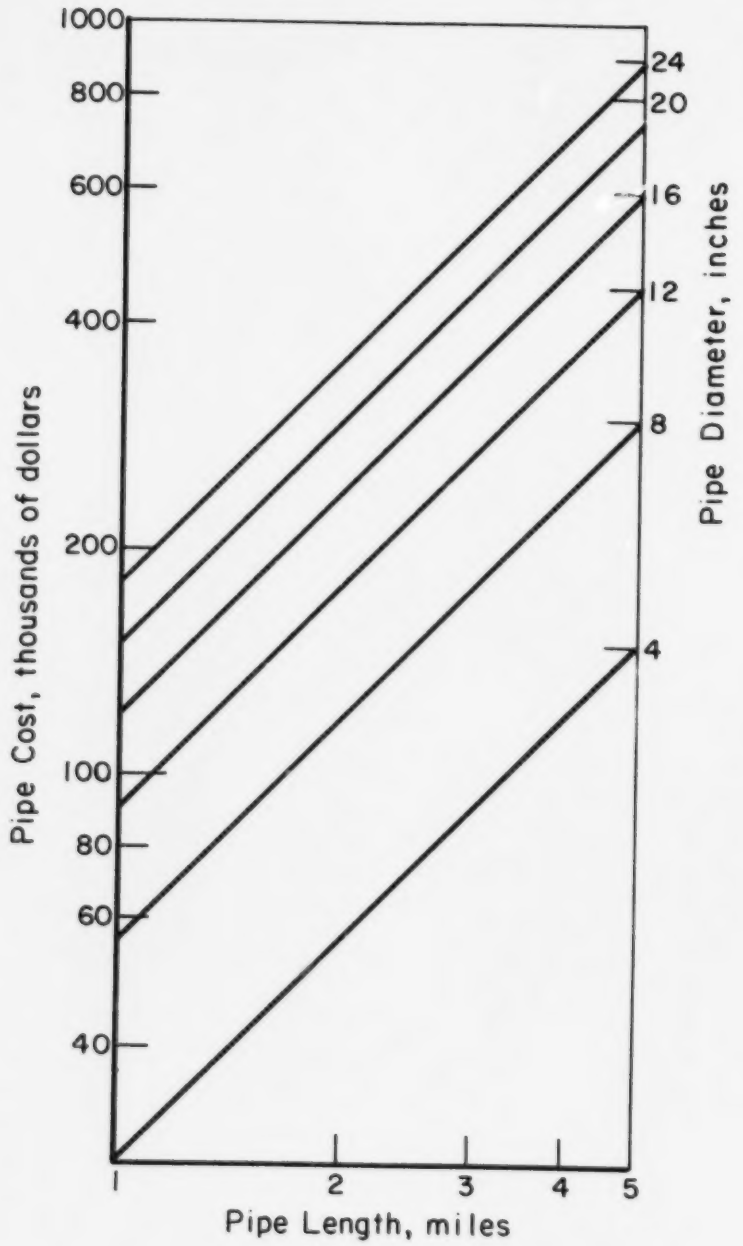

FIGURE 2. INSTALLED COST OF CARBON STEEL PIPE

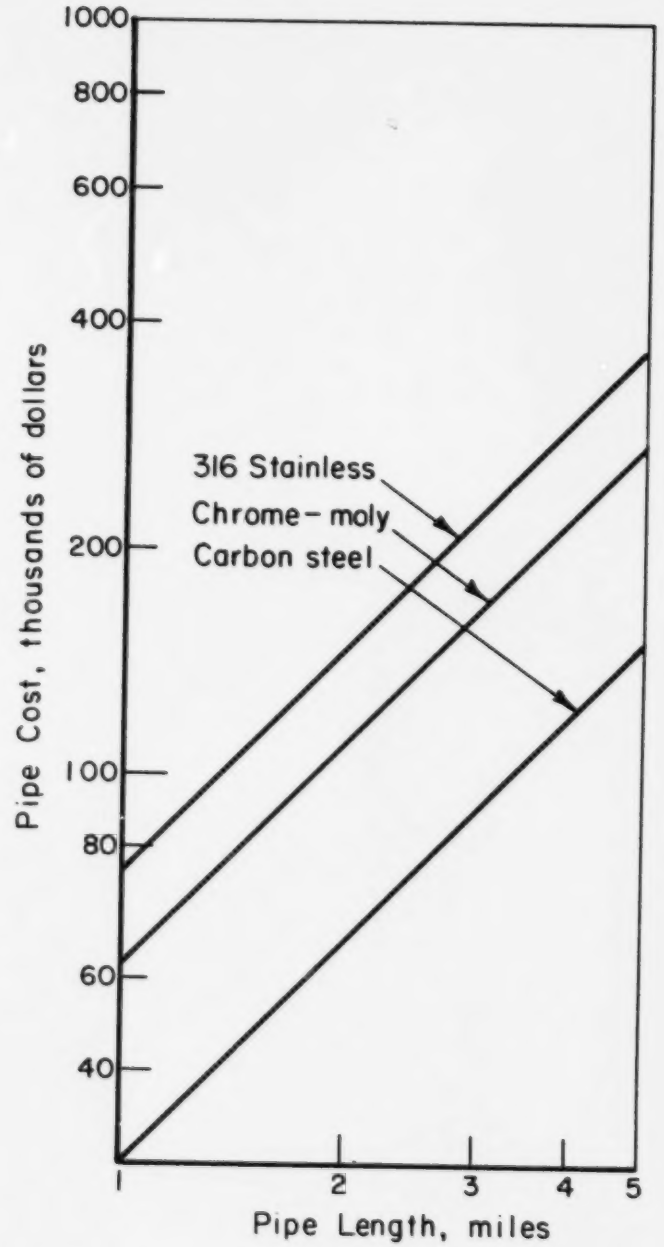

FIGURE 3. COMPARISON OF INSTALLED PIPE COST FOR THREE TYPES OF STEEL (FOUR-INCH PIPE LINE)

exceed the fan pressure rise. Two values for the maximum allowable pipe friction loss were selected and these were 20 and 35 inches of water gage. The higher value is about the highest practical value for centrifugal fans, considering that an additional 15 inches of water was assumed to be dissipated across the treatment equipment. The lower value of 20 inches of water was assumed for pipes about 1 mile long. If the higher value were used for the shortest pipes, a costly pressure loss would occur across the dampers in the pipes. For certain cases in relatively long pipes the air velocity exceeded 100 ft/sec for a certain pipe size; therefore, the next larger standard pipes were selected. Since the flow areas of the next larger standard pipes increase considerably for large pipes, the resulting pressure losses were below 20 inches of water, and these pipes were listed in the lower pressure-loss system. It is acknowledged that, once a system is selected, the pressure losses in all parts of the system must be compatible; but, it is also acknowledged that in an actual system all industrial plants might not be located the same distance from the centralized abatement plant as was assumed herein.

Tables 11 and 12 show the pipe material selected and the size and cost for a range of lengths with pressure losses of 20 and 35 inches of water gage, respectively. Only standard sizes of pipes were used. In Table 12 for the lime plant for the 4-mile length, for example, if 36-inch-diameter pipe were used, the pressure loss would have exceeded 35 inches of water gage. The next larger standard size of pipe is 42-inch diameter. The calculated pressure loss for this size was below 20 inches of water gage. If this pipe were installed in a system having 35 inches of water gage fan pressure, the 15 inches of water gage difference must be dissipated across a damper. Another alternative is to obtain specially manufactured pipe of the calculated diameter. Whether this would offset the operating costs of using a damper is beyond the scope of this report.

Finally, some consideration should be given to cooling of hot gas at the source before it enters the pipe line. There may be problems involved in sending 500 F gas through a buried pipeline because the ground temperature might eventually become excessively high above the line. On the other hand, if gases containing SO_2 and water are cooled below the dew point, which can be higher than 250 F depending on concentration, sulfurous and sulfuric acid may be formed. This might require a change in pipe material to stainless steel for corrosion resistance, and this would increase costs substantially. Final deliberations on the maximum allowable pipe temperature must take into consideration local codes, population density, and pipe location.

TABLE 11. PIPE COSTS FOR 20-INCH WATER-PRESSURE LOSS

		Pipe Length							
		1 Mile		2 Mile		4 Mile		5 Mile	
Industrial Plant	Pipe Material	Diameter, inches	Cost, million dollars	Diameter, inches	Cost, million dollars	Diameter, inches	Cost, million dollars	Diameter, inches	Cost, million dollars
Electric Power	Carbon Steel	48	0.360	48	0.720	--	--	--	--
Electric Steel Furnace	Carbon Steel	60	0.450	--	--	--	--	--	--
Gray-Iron Foundry	Carbon Steel	42	0.315	--	--	--	--	--	--
Lime	Carbon Steel	--	--	36	1.002	42	2.340	--	--
Sulfuric Acid	316 Stainless	36	0.671	--	--	--	--	48	4.470
Cement	Carbon Steel	78	0.585	84	1.260	--	--	--	--
Fertilizer	Carbon Steel	54	0.450	--	--	--	--	--	--
Totals			2.831		2.982		2.340		4.470

TABLE 12. PIPE COSTS FOR 35-INCH WATER-PRESSURE LOSS

Industrial Plant	Pipe Material	Pipe Length: 1 Mile Diameter, inches	1 Mile Cost, million dollars	2 Mile Diameter, inches	2 Mile Cost, million dollars	3 Mile Diameter, inches	3 Mile Cost, million dollars	4 Mile Diameter, inches	4 Mile Cost, million dollars	5 Mile Diameter, inches	5 Mile Cost, million dollars
Electric Power	Carbon Steel	--	--	--	--	54	1.215	60	1.800	60	2.250
Electric Steel Furnace	Carbon Steel	--	--	60	0.900	66	1.485	66	1.980	72	2.700
Gray-Iron Foundry	Carbon Steel	--	--	42	0.630	48	1.080	48	1.440	48	1.800
Lime	Carbon Steel	30	0.419	--	--	36	1.503	--	--	42	2.925
Sulfuric Acid	316 Stainless	--	--	36	1.342	42	2.346	42	3.128	--	--
Cement	Carbon Steel	--	--	--	--	78	1.755	84	2.520	90	3.375
Fertilizer	Carbon Steel	--	--	60	0.900	66	1.485	72	2.160	72	2.700
Totals			0.419		3.772		10.869		13.028		15.750

Fan Cost Estimates

Centrifugal fans[33] are recommended for transporting the polluted air to the processing plant. In sizing the fans, an additional pressure loss of 15 inches of water gage was added to that of the piping to account for the loss within the centralized processing plant. Therefore, the fan pressure rise was assumed to be either 35 or 50 inches of water gage. Fifty inches of water gage is near the limit of a simple, one-stage centrifugal fan. The total quantity of air handled from the seven industrial plants is 394,000 scfm (as calculated in Appendix C). Four fans of 110,000-scfm capacity each were considered in the estimates, but this does not provide standby capacity, if one of the blowers were shut down for maintenance.

The cost of four fans for 50 inches of water gage pressure is $252,000, and for 35 inches of water gage, $168,000. Each fan is a radial-blade, double-inlet centrifugal design. These estimates include cost of fans, inlet ducting, and installation costs. Costs of motors are shown separately below, since alternative prime movers might be substituted.

Motor Cost Estimates

Motors for the fans having 50 inches of water gage are each 2,000 hp, 1,200-rpm, 2,300-volt, 3-phase, 60-cycle units of open-type construction. Open motors were

selected, with a substantial cost reduction, on the assumption that the motors would be protected from the weather. Transformers for reducing line voltage of 13,200 volts to 2,300 volts are 2,000-kva capacity. Motor starters are included in the estimate.

The estimated cost of four motors, frames, starters, and transformers is $238,000. This estimate includes costs of material and installation.

On-site power generation should be considered. With electric power one is paying for transmission lines, power-generation apparatus, transformers, switchgear, and electric motors. The motors are fixed-speed units and require either throttling for flow control or variable-speed (fluidic) fan couplings. Use of on-site gas-turbine or gas-engine fan drives would reduce considerably both the investment cost and the operating cost of the system. With direct drive from engines or turbines, fan speed could be adjusted to match loads. In this regard, the fan power for the illustrative system, at 8,000-hp, would absorb about 1/4 of the power output of the 25,000-kw power plant in the system.

Damper Cost Estimates

Dampers in each of the pipelines are necessary to maintain a constant pressure drop in each of the pipes of the transportation system. As the effluent-gas rates are not constant at all times, pressure controllers are included.

An estimate of the costs of the damper system was made based on quotations by Hayes who cited a cost of $4500 plus 500 installation for a 48-inch butterfly gas valve made of carbon steel.[34] The following assumptions were made in calculating the required costs:

(1) Valve costs were proportional to the area of the pipe

(2) Installation costs were proportional to the diameter of pipe

(3) Valves made from alloy steels would be five times the cost of those from carbon steel

(4) The pressure controller for each damper would cost $10,000.

Calculations for damper costs are shown in Appendix B. Capital costs were estimated to be $141,000. Operating cost would consist only of maintenance, which is usually relatively high for control devices. It was estimated at 5 percent of the capital cost per year, that is $7,050 per year.

Operating Costs

Electricity and maintenance are the major operating costs, along with the 20 percent financial charge. Using a value of $0.01 per kwhr for electric power, the cost of running the four 2000-hp motors would be $523,000 per year.

The cost of pipeline maintenance, i.e., service to cathodic protectors, occasional cleaning, etc., and of motor service was roughly estimated at 1 percent of the capital expenditures over a 10-year period, or $11,000 per year. Thus total operating expense for a 3-mile distance was $523,000 + $11,000 + $7,000 + 20 percent of $11,500,000 = $2,841,000.

Total Gas-Handling Cost Estimates

To obtain a total cost estimate for the gas-handling system, the pipe costs of Tables 11 and 12 were added to the fan and motor cost, and the damper cost. These costs have been plotted in Figure 4. The cost of the pipelines is about 90 percent of the total cost of the gas-transportation system. Calculations are shown in Appendix B.

Summary of Major Assumptions and Specifications

(1) All plants are located equidistant from the centralized treatment facility.

(2) The treatment plant reduces particulate matter and corrosive gases below a limit which would be harmful to the fans.

(3) The fans are located after the pollution-control device.

(4) The fans can develop 50 inches of water suction; 35 inches through the pipelines, and 15 inches across the control equipment.

(5) Pipes are to be buried 4 to 5 feet underground in industrial locations.

(6) The material of construction for the pipelines is carbon steel, except for the lime plant and sulfuric acid plant which utilize molychrome and 316 stainless, respectively.

EMISSION CONTROL FOR TOTAL GAS STREAM

This section deals with the central treatment plant, whose function is to treat the mixed effluents from the seven individual plants in one large control device.

Control Device

The compositions of the mixture of gas streams from the seven plants is shown in Table 13. Calculations for this table are shown in Appendix C. It can be seen that the gas stream contains various compounds which will react with each other, mainly

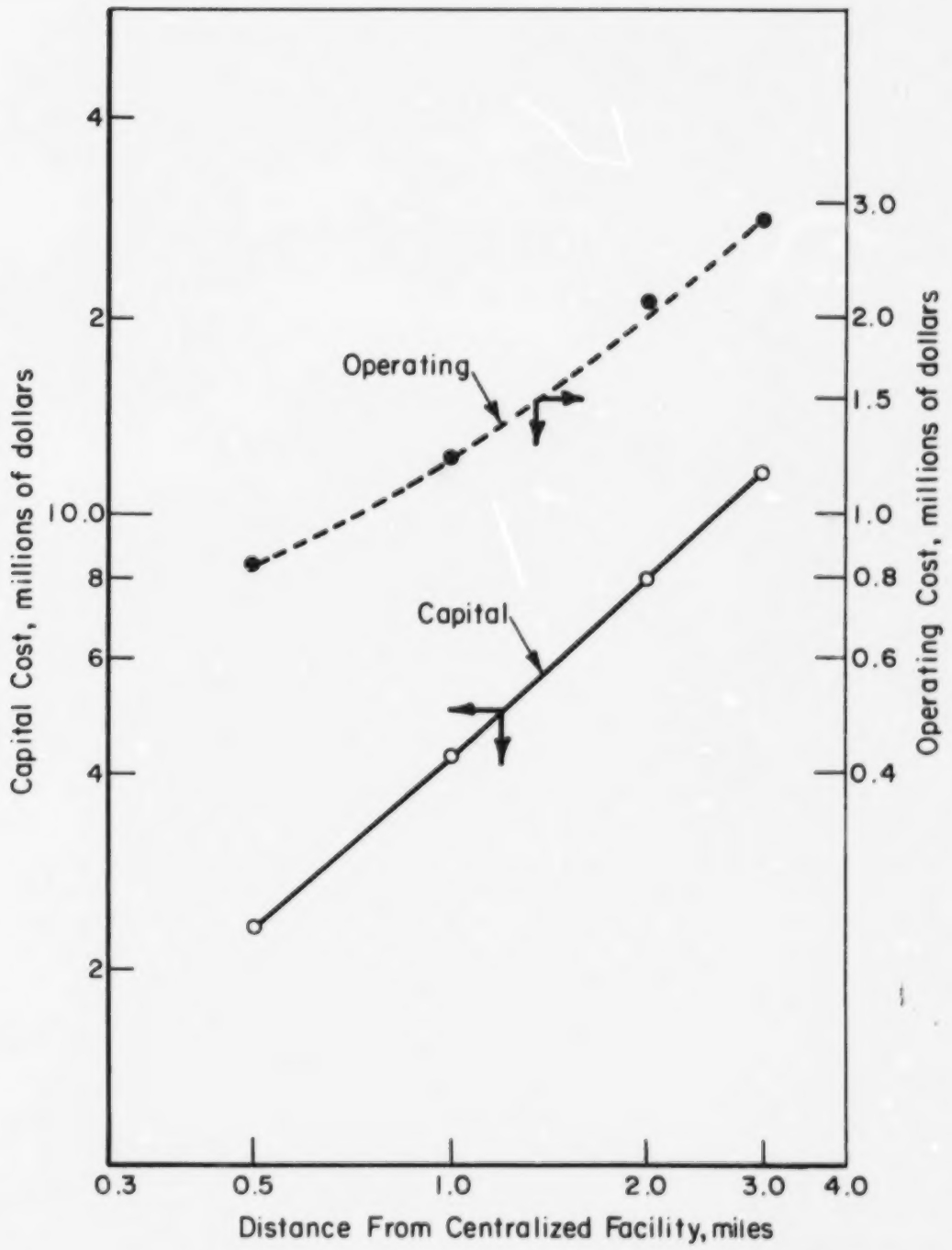

FIGURE 4. INSTALLED COST OF GAS TRANSPORT SYSTEM

acidic compounds such as SO_2, sulfuric acid, and CO_2, and basic constituents from the cement and lime plants. By a consideration of the free energy changes accompanying the various reactions (shown in Appendix C), it was shown that the most favored reaction would be the reaction of lime and SO_2. However, the level of various reactive acidic components in the gas after mixing would depend on factors such as (1) the amount of lime present for reaction, (2) temperature of the gas, (3) reactivity with cement, and (4) residence time of the reactants.

Thus, it is impossible to give concentration levels of SO_2, sulfuric acid, and the fluorides after the gases are mixed. In the present case, there would not be enough CaO in the emissions to remove the acidic pollutants entirely, necessitating their removal by the control device. It is felt that since sulfuric acid, the fluorides, and SO_2 are relatively soluble in water, and react easily in an aqueous system, a wet-scrubbing device would give reasonable control. The water used in the wet scrubber should have lime dissolved in it in order to form $CaSO_4$ and CaF_2. These two compounds, because of their low solubility in water, could be removed by precipitation and filtration, along with the dust emissions.

TABLE 13. MIXED GAS COMPOSITION

Variable	Before Scrubbing	After Scrubbing
Gas Flow, 1000 acfm	627	530
Gas Temperature, F	320	240
Particulate Loading, gr/ft^3	2.7	<.01
Gas Composition, volume percent		
N_2	69.3	67.0
O_2	9.2	8.8
CO_2	10.4	10.0
H_2O	10.7	13.2
SO_2	.04	--
CO	1.1	1.0
Fluorides	Trace	--
H_2SO_4	Trace	--

Cost Considerations

Chemical Construction Corporation supplied necessary information on types of wet scrubbers that would be suitable for handling the emissions encountered.[12] They felt that a high-energy venturi scrubber would accomplish the desired degree of control, and they had recent data on a scrubber installation for a power plant.

The scrubber for that power plant cost $750,000, including installation, and it handled 420,000 cfm at 420 F. Since this was one of the largest they manufactured, costs for this study were scaled up on a cfm basis, making the capital cost of the scrubber for the emissions in the present case equal to

$$\frac{627,000}{420,000} \times \frac{880}{780} \times \$750,000 = \$1.27 \text{ million.}$$

Unfortunately, no operating cost data were available. To estimate these costs, the results of Sargent[17] were scaled up to the present case, the cost of lime necessary to remove 50 percent of the acid gas pollutants (the rest assumed to react with the lime-plant emissions) was added. Lime cost used was $15.50 per ton (Oil, Paint and Drug Reporter, December, 1969). Detailed cost calculations can be found in Appendix B.

DISCUSSION

A detailed cost summary is shown in Table 14. Total cost of pollution control on an individual basis is about $5.4 million capital, and $1.7 million annual operating. In comparison, the costs for centralized control (3-mile radius) would be $14.0 million capital and $3.6 million annual operating.

TABLE 14. COST SUMMARY

Plant	Equipment		Stack		Buildings, etc.		Labor, dollars/ year	Total Capital, dollars	Total Operating, dollars/ year
	Capital, dollars	Operating, dollars/ year	Capital, dollars	Operating, dollars/ year	Capital, dollars	Operating, dollars/ year			
Cement	745,000	206,000	121,800	24,400	651,000	130,200	19,700	1,517,800	380,300
Lime	50,800	16,000	77,800	14,600	97,000	19,400	19,700	225,600	69,700
Sulfuric Acid	55,000	11,000	77,800	14,600	101,000	20,200	19,700	233,800	65,500
Power	645,000	250,000	77,800	14,600	542,000	108,400	19,700	1,264,800	392,700
Fertilizer	210,000	99,800	100,000	20,000	233,000	46,600	19,700	543,000	186,100
Gray-Iron	400,000	168,000	77,800	14,600	358,000	71,600	19,700	835,800	273,900
Electric Arc	343,000	195,000	100,000	20,000	332,000	66,400	19,700	775,000	301,100
Total	2,448,800	945,800	633,000	122,800	2,814,000	462,800	137,900	5,395,800	1,669,300
Centralized Control	1,270,000	539,000	177,000	35,400	1,083,000	216,600	78,800	2,530,000	869,800
Gas Transport (3 miles)	11,500,000	2,841,000	--	--	--	--	--	11,500,000	2,841,000

It is noteworthy that the premise of one large treatment plant being more economical than several smaller treatment plants is verified; the central pollution-control plant costs $2.5 million in comparison to a $5.4 million cost for individual control facilities. However, the cost of transporting the effluents to the centralized plant is very high; being $11.5 million for the 3-mile radius. As the transportation cost is a function of distance, the difference between the cost of centralized and individual pollution control can be plotted versus distance of the plants from the treatment facility to determine a "break-even" point. Figure 5 shows that if the individual plants were located about 1/2 mile from the centralized treatment facility, it would appear more economical to treat the gases in the central plant.

Before drawing definite conclusions, however, it is important to consider several factors that have a bearing on the feasibility of the centralized concept.

(1) The cost of waste disposal was not estimated as part of the economics because the disposal cost is a function of many variables, most of which can only be guessed. However, valid observations about the effect of waste disposal on the economics of the centralized concept can be made if one assumes that the disposal cost per ton of material is the same for both the centralized plant and the individual plants. This assumption is not unreasonable because the major variables affecting this cost, such as location, labor costs, proximity of disposal facilities and type of waste, should be the same in both cases.

    If none of the effluents have any value to the individual plants, that is, the effluents cannot be recycled or sold as product, then the disposal costs should not influence the economics significantly because approximately the same amount will have to be disposed of in both the centralized facility and in the individual plants. However, in general, this is not the case. The sulfuric acid mist can be collected as a saleable product if individual control were used. For the case of centralized control, mixing of the effluents from the seven plants would render the sulfuric acid valueless because of its reaction with the lime either from the lime plant emissions or in the scrubber. For the sulfuric acid plant used in this study, H_2SO_4 emissions amount to about $25,000 per year ($H_2SO_4$ at $25/ton).

    The only other emissions which may be of value are those from the lime and cement plants. For the value of these emissions to be realized, however, they must be returned to the kiln, a practice which is not in general use at the present time because of the complexities involved.

(2) Burial of the hot pipes would result in localized heating of the ground above the pipes. This would probably eliminate any vegetation in this area. This may be undesirable from an esthetic viewpoint and would increase run-off somewhat.

(3) A malfunction in the centralized pollution-control facility could cause serious problems. If the operation of the scrubber were to be discontinued for repair, a relatively large amount of pollutants would be

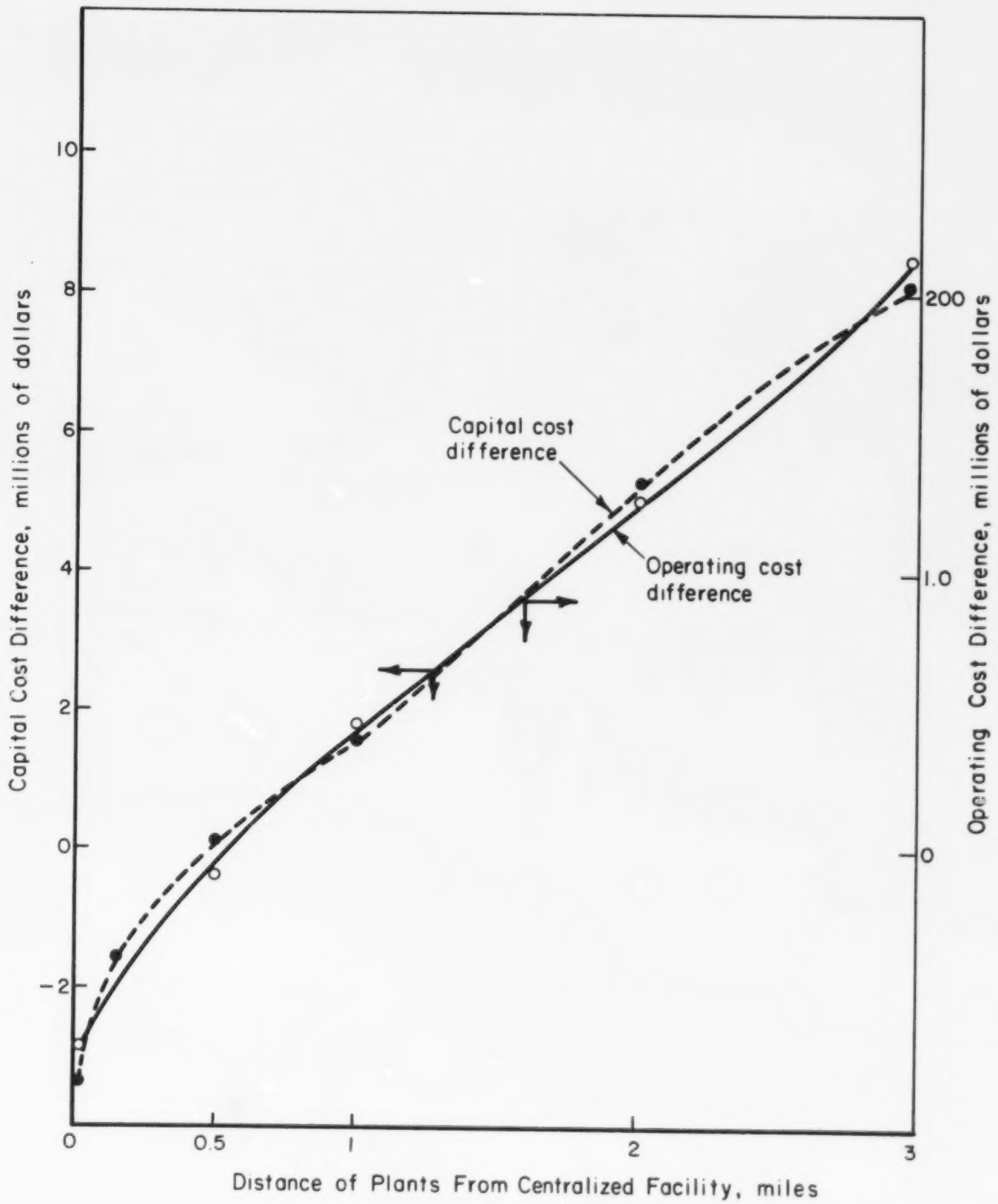

FIGURE 5. EFFECT OF DISTANCE ON COST DIFFERENCE BETWEEN CENTRALIZED AND INDIVIDUAL CONTROL ABATEMENT

released over a small area. This situation could possibly cause serious temporary problems; whereas, in the case of individual control, the chance of all seven pollution devices malfunctioning at the same time is very remote. Problems of this nature could be alleviated somewhat by utilizing two scrubbers in parallel, each handling one-half of the effluents, at an increase in cost.

(4) If a malfunction occurred in the transportation system, i.e., failure of the pipeline or fans, it would be necessary to vent the emissions elsewhere. This situation might be impossible for a given plant. Therefore, some extra capacity in the transportation system would probably have to be provided; for example, an extra line and extra fans. To provide for this contingency, additional expense would be involved.

Some information pertinent to this study was obtained from Brink[18] during discussion of the demister control units. He stated that Monsanto had done a centralized air-pollution study for one of their own plant complexes similar to the one presented in this report. On the basis of results of their study, Monsanto rejected the centralized concept as being both too expensive and presenting too many technical difficulties. It is interesting to note that the major factor that determined the cost of centralized treatment in the Monsanto study was also the cost of gas transportation.

CONCLUSIONS

On the basis of this study of a centralized-air-pollution-abatement concept, it is concluded that the centralized concept is not economically feasible, mainly because the cost of transporting the gas to the centralized facility is so high that the individual plants would have to be less than about 1/2 mile from the central treatment plant for any economic gain over individual treatment to be realized. From a practical standpoint, it would be difficult to envision seven plants so close together because of their individual land requirements.

Considerations of waste disposal, natural esthetics, and potential emergency problems associated with the failure of the centralized equipment all tend to reduce the attractiveness of the concept further.

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APPENDIX A

GAS COMPOSITIONS

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APPENDIX A

GAS COMPOSITIONS

Cement Plant Calculations

Gas Composition

Usual ratio of $CaCO_3$ to coal is 4 to 1.[35] Reactions are

$$C + O_2 \longrightarrow CO_2 \qquad (1)$$

$$CaCO_3 \longrightarrow CO_2 + CaO \qquad (2)$$

Assume 10 percent excess air:

oxygen required per lb of C [Equation (1)]

$$= \frac{32}{12} \times 1.1 = 2.94 \text{ lb } O_2 = 0.0918 \text{ moles } O_2 \ .$$

Therefore, nitrogen $= \frac{0.0918}{0.21} \times 0.79 = 0.345$ moles N_2 ,

oxygen in stack $= 0.1 \times 0.0918 = 0.00918$ moles O_2 ,

and carbon dioxide:

From Equation (1): $= \frac{44}{12} = 3.67$ lb

From Equation (2): $= \frac{44}{80} \times 4 = 2.20$ lb

Total CO_2 $= 5.87$ lb or 0.133 moles CO_2 .

On a dry basis:

$$CO_2 = \frac{0.133}{0.478} = 28 \text{ percent}$$

$$O_2 = \frac{0.00918}{0.478} = 1.9 \text{ percent}$$

$$N_2 = \frac{0.345}{0.478} = 70 \text{ percent} \ .$$

On a wet basis:

From Kreichelt[1], water content = 31 percent

CO_2 = 0.69 x 28 = 19.5 percent

N_2 = 0.69 x 70 = 48.2 percent .

Lime Plant Calculations

Gas Composition

From Lewis and Crocker[2], CO_2 = 31.8 percent for a 4:1 lime-to-fuel ratio.

$$C + O_2 \longrightarrow CO_2 \quad (3)$$

$$CaCO_3 \longrightarrow CO_2 + CaO \quad (4)$$

On a basis of 1 lb of C and 4 lb of CaO:

moles of CO_2 from Equation (3) = 0.0833

moles of CO_2 from Equation (4) = 0.1111

Total = 0.1943 .

Thus, other constituents = $\frac{0.1943}{0.318}$ x 0.682 = 0.416 moles .

Assume 10 percent excess air, then

O_2 = 0.0833 x 1.1 = 0.0916 moles

O_2 in stack = 0.0083 moles

$N_2 = \frac{0.0916}{0.21}$ x 0.79 = 0.345 moles

moles of water and SO_2 = 0.416 - 0.353 = 0.063 moles

assume SO_2 = 0.1 percent = 0.001 x (0.416 + 0.194)

= 0.0006 moles.

Thus, water = 0.062 moles

N_2 = 0.345/0.610 = 56.5

CO_2 = 31.8

O_2 = 0.008/.610 = 1.4

$H_2O = 0.062/.610 = 10.3$

$SO_2 = 0.1$.

Sulfuric Acid Plant Calculations

Exit-Gas Composition

From Cuffe and Dean[5], the average SO_2 concentration entering the converter is 8.3 percent, and SO_2 in stack is 0.26 percent.

$$S + O_2 \longrightarrow SO_2 \ .$$

To calculate gases input (basis 1 mole of S): Let x = moles of excess O_2 input.

input O_2 = 1 + x

input N_2 = (1 + x) 0.79/0.21

output O_2 = x

output SO_2 = 1

Total moles entering converter = (1 + x) 0.79/0.21 + 1 + x

= 4.76 + 4.76x .

Solving: $\frac{1}{4.76 + 4.76x} = 0.083$

x = 1.52 moles of O_2 excess .

To calculate SO_2 in stack; the moles of gases for the stack are as follows:

N_2 = 9.45

O_2 = 1.52

$SO_2 = \frac{x}{10.97 + x}$.

Know that $x/(10.97 + x) = 0.0026$. Thus, $x = 0.0029$.

N_2 = 9.45/11 = 85.8 percent

O_2 = 1.52/11 = 13.8 percent

SO_2 = 0.26 percent

SO_3 = 0.03 percent .

Gray-Iron Foundry Calculations

The calculated example in Reference (13), pages 292-296, was used as a basis for gaseous compositions. This reference gives weight of emissions for a melting rate of 21.91 tons of iron changed per hour. The gas rates are given in lb/hr at 850 F.

CO_2 = 6,780 lb/hr = 154 moles/hr

CO = 4,370 lb/hr = 153 moles/hr

N_2 = 24,500 lb/hr = 875 moles/hr

H_2 = 26.3 lb/hr = 13.2 moles/hr .

To recover some of the sensible heat from the stack gases, a widely used and successful method is the Griffin system. Also in this system, provision is made for introducing air, which burns the CO to CO_2, and the H_2 to H_2O, and recovers the heat of combustion. It will be assumed that this system will be used. The reactions are:

$$CO + 1/2\ O_2 = CO_2$$

$$H_2 + 1/2\ O_2 = H_2O \ .$$

Moles of O_2 required = 1/2 (153 + 13.2) = 83.1 moles/hr.

Assuming 10 percent excess air, oxygen leaving stack = 8.31 moles/hr.

N_2 = 875 + 79/21 (91.4) = 1219 moles/hr.

CO_2 = 154 + 153 = 307 moles/hr.

The temperature leaving the heating chamber in the Griffin system is about 500 F. Therefore, to convert from moles/hr to acfm, the factor $\frac{359}{492} \times \frac{960}{60} = 11.65$ is used.

N_2 = 1219 moles/hr = 14,200 acfm

CO_2 = 307 moles/hr = 3,580 acfm

O_2 = 8.3 moles/hr = 100 acfm

H_2O = 26.3 moles/hr = 300 acfm

Total = 18,180 acfm .

From these numbers, stack-gas composition is easily calculated.

An average cupola operation appears to utilize three cupolas, melting 60 tons of iron per hour. Therefore, emissions would be

18,180 x 60/21.91 = 49,600 acfm .

Electric Furnace

For the electric furnace, both CO and CO_2 are produced in the reaction of the carbon electrodes with air. Information necessary for the calculation of the composition of the stack gases was obtained from Barnes and Lownie[16].

The two reactions occurring are:

$$C + 1/2\ O_2 \longrightarrow CO \qquad (5)$$

$$CO + 1/2\ O_2 \longrightarrow CO_2 \qquad (6)$$

The following assumptions were made: (1) Equation (5) takes place with stoichiometric amount of O_2, (2) Equation (6) takes place with 500 percent excess air, (3) the final CO_2/CO ratio is 5.

Basis: 1 mole of C

Let x = number of moles of CO reacted in Equation (6).

Then, since 1 mole of CO is produced by Equation (1), there will be (1-x) moles of CO in the exit gas stream.

Thus, $x/(1-x) = 5$ and $x = 5/6$.

O_2 inlet	$= 0.5 + 1/2\ (5/6) \times 5$	$= 2.58$ moles
O_2 used up	$= 0.5 + (1/2)\ (5/6)$	$= 0.92$ moles
O_2 in stack	$= 2.58 - 0.92$	$= 1.66$
N_2 in stack	$= (2.58 \times 79)/21$	$= 9.7$
CO_2 in stack		$= 0.833$
CO in stack		$= 0.166$

From these results, the gas composition follows readily.

APPENDIX B

COST CALCULATIONS

APPENDIX B

COST CALCULATIONS

Cement Plant

Electrostatic precipitator efficiencies for various particle sizes are given by Sargent[17] (Table V), and approximate screen analysis of the cement dust by Kreichelt, et al.[1]. Total efficiency calculations are tabulated below:

Particle Size, microns	Amount Present, weight percent	Precipitator Efficiency, percent	Dust Collected, percent of total
2.5	22.5	77.0	17.3
5	7.5	90.5	6.8
10	12.5	95.0	11.9
20	17.5	96.0	16.8
30	20.0	96.5	19.8
40	10.0	96.8	9.7
50-60	10.0	97.7	9.8
		Total	91.6

Data from Barnes and Lownie[16] indicate that about a 10 percent increase in capital and operating costs is necessary to decrease the outlet dust loading by 2.5 times. The amount of decrease necessary for the present case would be

$$\frac{(1 - 0.916) \times 5.4}{0.05} = 9.14 \quad .$$

Thus, since $(2.5)^{2.4} = 9.14$, the percentage increase in cost required would be $2.4 \times 10 = 24$ percent.

Size Considerations

The data in Sargent show a cost of the electrostatic precipitator to be $233,000 for a gas flow rate of 60,000 cfm at 68 F. At this temperature, the flow rate of cement-plant emissions would be $193,000 \times 528/864 = 118,000$ cfm.

To scale up the cost data of Sargent, use the relationship

$$\frac{\text{Cost}_1}{\text{Cost}_2} = \left[\frac{\text{flow}_1}{\text{flow}_2}\right]^m \quad .$$

The index "m" was calculated to be 0.7 from data of Barnes and Lownie (page C-42).

The cost of a 118,000-cfm unit would be

$$233,000 \times \left[\frac{118,000}{60,000}\right]^{0.7} = \$375,000 \quad .$$

Addition of the 24 percent for the required efficiency increase, and adjustment of cost to 1969 levels would result in a capital investment of $490,000.

Yearly operating costs would come to about

$$1.24 \times (\$2,000 + \$1,300) \times \left[\frac{118,000}{60,000}\right] + 20\% \text{ of } \$490,000$$

$$= \$8,000 + \$98,000 = \$106,000 \quad .$$

Costs for electrostatic precipitation have been reported by Barnes and Lownie and can be calculated for this case:

$$\text{Capital Cost} = \$905,000 \times \left[\frac{193,000}{185,000} \times \frac{960}{864}\right]^{0.7}$$

$$= \$1,000,000 \quad .$$

Similarily, yearly operation costs = $306,000

Averaging the two costs from Sargent and Barnes and Lownie gives

Capital cost = $745,000

Operating cost = $206,000/year

Fertilizer Plant

Barnes and Lownie[16] report costs of a low-energy venturi-type wet scrubber, handling 55,000 cfm at 70 F, to be $179,000 and operating costs to be about $75,000 yearly (Page C-26). To scale these prices to the fertilizer flow rates (100,000 cfm at 90 F), it was assumed that the cost index was 0.7.

Thus, control cost for the fertilizer plant would be

$$\$179,000 \times \left[\frac{100,000}{55,000} \times \frac{530}{550}\right]^{0.7} = \$265,000 \quad .$$

The operating costs were assumed to be proportional to the volume of gases treated, with the exception of the 20 percent capital charges. Therefore, operating cost per year for the fertilizer-plant control would be

$$\frac{100,000}{55,000} \times \frac{530}{550} \ (\$22,000 + \$7,000 + \$10,000) + 20\% \text{ of } \$265,000$$

$$= \$63,000 + 52,400 = \$115,400/\text{year} \quad .$$

In a similar manner, costs for the treatment of fertilizer-plant gas can be calculated from the data of Sargent[17] (Table VI),

$$\text{Capital cost} = 107,000 \times 1.05 \times \left[\frac{100,000}{60,000} \times \frac{528}{550}\right]^{0.7}$$

$$= \$155,000$$

$$\text{Operating cost} = \frac{100,000}{60,000} \times \frac{528}{550} \times (\$12,100 + \$1,000 + \$18,000) \times 1.05$$

$$+ 20\% \text{ of } \$155,000$$

$$= \$53,200 + \$31,000 = \$84,200/\text{year}$$

The estimate of costs was obtained by averaging the costs from both sources.

$$\text{Capital cost} = (\$265,000 + \$155,000)/2 = \$210,000$$

$$\text{Operating cost} = (\$115,400 + \$84,200)/2 = \$99,800/\text{year}$$

Dampers

The table below details the cost calculations for the dampers required. This table is based on a cost of $4500 for a carbon steel valve of 48-inch diameter, plus $500 installation charge.

Plant	Pipe Diameter, inches	Valve Cost, dollars	Installation, dollars
Cement	78	11,600	800
Lime	36	12,700	370
Sulfuric Acid	42	17,200	430
Power	54	5,700	560
Fertilizer	60	7,000	625
Gray-Iron	48	4,500	500
Electric Furnace	66	8,500	690
	Totals	67,200	3,985

Assuming $10,000 for pressure controllers for each pipeline, total capital cost would be

$$\$67,200 + \$3985 + 7 \times \$10,000 = \$141,000$$

Operating cost = 5 percent of $141,000 + 20 percent of $141,000 = $35,300/year

Scrubber Operating Costs

Data of Sargent[17] can be scaled up as follows:

$$\text{Cost} = \frac{627,000}{60,000} \times \frac{528}{780} \times (\$18,820 + \$12,100 + \$1,000)$$

$$= \$225,000/\text{year} \quad .$$

The cost of the lime required is:

$$\frac{.527}{2} \times \frac{56 \times 60 \times 24 \times 365}{2000} \times \$15.50 = \$60,000/\text{year} \quad .$$

Operating cost of high-energy wet scrubber =

$$\$225,000 + \$60,000 + .20 \times \$1,270,000 = \$539,000/\text{year} \quad .$$

Total Gas Handling

3-Mile Distance

Capital cost = pipeline cost + fan cost + motor cost + damper cost

= \$10,869,000 + \$252,000 + \$238,000 + \$141,000

= \$11,500,000

Operating cost consisted of electricity, pipeline and damper maintenance plus the 20 percent of capital investment.

Operating cost = \$523,000 + \$11,000 + \$7,000 + 20% of \$11,500,000
= \$2,841,000/year

2-Mile Distance

It was assumed that the same size pipes, fans, motor, and dampers would be used in this case as for the 3-mile case.

Thus,

Capital cost = 2/3 x \$10,869,000 + \$252,000 + \$238,000 + \$141,000

= \$7,881,000

Operating cost = $523,000 + $11,000 + $7,000
+ 20% of $7,881,000

= $2,114,000/year

1-Mile Distance

Pipeline cost (Table 11) = $2,831,000 + $1,002,000/2

= $3,332,000

Fan Cost = $168,000

Motor Cost = $119,000

Damper Cost = $100,000

Total Capital Cost = $3,700,000

Operating costs:

Electricity = $523,000 x 35/50 = $367,000

Pipeline maintenance = 0.1 percent of $3,833,000

= $3,800

Damper maintenance = $5,000

Total operating costs = $376,000 + 20 percent of $3.7 million

= $1,116,000/year

1/2-Mile Distance

Again, it was assumed that the same size pipes, etc., would be used in this case, as for the 1-mile case.

Capital cost = 1/2 x $3,833,000 + $168,000
+ $119,000 + 100,000

= $2,293,000

Operating cost = $376,000 + 20 percent of $2,293,000

= $834,600/year

Figure 5 shows the difference between the costs of centralized pollution control and individual control. For example, for a 3-mile distance, centralized control would cost

$11,500 + $2,530,000 = $14,030,000 capital

and $2,841,000 + $869,000 = $3,710,000/year operating

The difference would be

$14,030,000 - $5,396,000 = $8,634,000 capital

and $3,710,000 - $1,669,000 = $2,041,000 operating.

Pipe Line

O'Connor[27] shows the cost per mile of a 26-inch-diameter pipe is $195,000 for laying pipe in U. S. Gulf Coast marsh land where costs can be expected to be higher than normal. This is about $30,000 per mile for 4-inch-diameter pipe assuming a linear relationship between cost and pipe diameter. Pipeline Industry[24] shows an extensive amount of pipe cost data. Reducing these data on the same basis, using only pipe costs with no substations, the cost per mile of 4-inch-diameter pipe is about $26,000 to $30,000 and the relationship between pipe cost and diameter up to 42 inches in diameter is nearly linear. Stark[26] shows curves for 4, 6, and 8-inch diameter pipe. The 4-inch-diameter pipe cost is $26,400 per mile and a nearly linear relationship between cost and diameter is shown. These references do not break the pipe costs into separate items, but they do discuss a few variables which can vary the cost.

Guthrie[25] shows a detailed cost breakdown, but costs are nearly double those shown above and were not used directly. These costs were high for several reasons. One reason is that they are estimates rather than actual costs, and Guthrie admits that these cost estimates can decrease by as much as 40 percent under competitive bidding. Another reason is that these estimates were for short lengths of line such as those run in an industrial plant rather than cross country.

Based on these data it was believed that the installed cost of $30,000 per mile represents a reasonable estimate for laying several miles of pipe.

White[23] shows a somewhat detailed cost breakdown for laying eight sizes of pipe. Costs per foot of 32-inch-diameter pipe shown by White were divided by eight to obtain costs on the basis of a 4-inch-diameter pipe and multiplied by 5280 ft/mile to obtain the costs shown below:

Estimated Cost of Installed 4-Inch-Diameter Pipe	Cost per Mile	
Item – Pipe, Grade X-52, 0.312-inch wall	$7,410	
Coating material	376	
Cathodic protection ($0.02/ft, regardless of pipe size)	106	
Total pipe material cost		$7,892

Surveying, mapping, pipe X-ray (0.05/ft)	$ 264	
Damages	2,400	
Freight	340	
Construction	4,650	
Total construction and closely allied services		$ 7,654
Total material and construction		$15,546

Subtracting this cost from the assumed $30,000 per mile cost, the remaining cost of $14,454 includes land easement rights, overhead, etc., not included in the above tabulation.

Another estimate was made using formulas shown by Guthrie and on the basis of costs by White, since it is believed that material costs by Guthrie might be excessively high for the air pollution abatement application. This estimate is as follows:

Item	Cost per Mile, dollars
Total pipe material costs, including X-ray	8,156
Indirect costs – 1.34 x material cost	10,920

Breakdown of 1.34 factor:

Item	Percent of Material Costs
Material	100.0
Engineering[a]	14.8
Direct labor	78.0
Construction Overhead[b]	38.4
Sales tax	3.0
Freight	5.0
Total material and O. H.	239.2

$$\frac{\text{Total material + Overhead}}{\text{Material + Direct Labor}} = \frac{239.2}{178.0} = 1.34$$

	19,076
Contingency (10 percent of 19,076)	1,908
	$20,984

(a) This cost includes direct engineering costs – pipe-circuit analysis, analytical engineering, plot plans – and indirect office costs, burden and overhead, and contractor fees.

(b) This cost is 49.2 percent of direct labor which equals 38.4 percent of material and includes field labor benefits and statutory burdens, field supervision, some average rigging, equipment rental, small tools, etc.

A detailed cost of land and associated costs were not available, but approximately $9,000 per mile for the basic 4-inch-diameter pipe was used in the estimate reported.

Obviously, if land need not be purchased or leased, the installed pipe cost might decrease by about one-third of the former estimate of $30,000 per mile for 4-inch-diameter pipe.

Sulfuric Acid

Figure B-1 shows cost versus cubic feet per minute for demisters of two efficiencies.

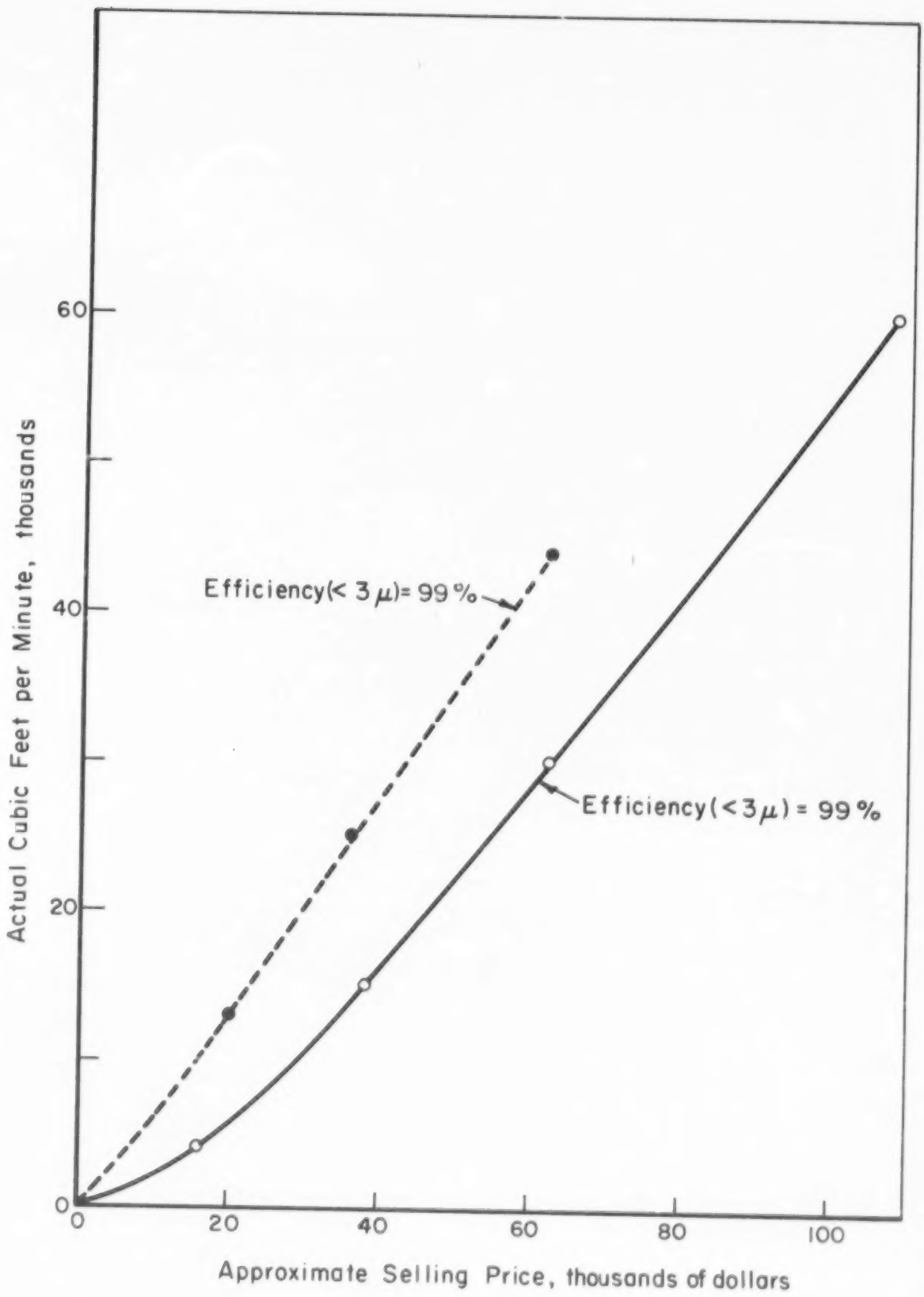


FIGURE B-1. PRICES FOR BRINK DEMISTERS

APPENDIX C

TOTAL GAS STREAM CALCULATIONS

APPENDIX C

TOTAL GAS STREAM CALCULATIONS

Gaseous Composition

The amounts of various constituents for the mixed streams of gases are shown in Table C-1. From the totals, gas compositions of the mixed stream can be calculated.

TABLE C-1. GAS FLOW RATES FOR INDIVIDUAL PROCESSES

Process	Gas Flow, 1000 acfm	Temp, F	Component, moles/minute					Particle, lb/min	Other
			N_2	O_2	CO_2	H_2O	SO_2		
Cement	193.0	404	148	4.0	59.8	95.0	--	149	--
Lime	29.0	910	17.1	0.4	9.7	3.1	0.0303	24.9	--
Sulfuric Acid	26.5	154	50.6	8.2	--	--	0.1535	--	Acid, 0.49 lb/min
Power	70.0	300	96.4	4.4	18.0	7.6	0.253	15.0	--
Fertilizer	100	90	196	52	--	--	--	--	SiF_4 and HF 2.1 lb/min
Gray Iron	496	500	55.3	0.3	13.9	1.2	--	14.2	--
Electric Arc	116.0	500	130.0	22.3	2.2	--	--	41.4	CO, 11.1 mole/min
Totals			693.4	91.6	103.6	106.9	0.437	244.5	CO: 11.1

Reaction Calculations

The gaseous emissions from the seven plants contain various acid gases and basic solids which can react together. Some reactions that may occur (along with the standard free energy change, ΔF°) are as follows:

$$CaO + SO_2 + 1/2\,O_2 \longrightarrow CaSO_4, \quad \Delta F^\circ = -96 \text{ kcal/mole}$$

$$CaO + CO_2 \longrightarrow CaCO_3, \quad \Delta F^\circ = -33 \text{ kcal/mole}$$

$$SiF_4 + 2CaO \longrightarrow 2CaF_2 + SiO_2, \quad \Delta F^\circ = -70 \text{ kcal/mole}$$

$$H_2SO_4 + CaO \longrightarrow CaSO_4 + H_2O, \quad \Delta F^\circ = -29 \text{ kcal/mole}$$

$$2HF + CaO \longrightarrow CaF_2 + H_2O, \quad \Delta F^\circ = -27 \text{ kcal/mole}$$

All the reactions have negative free energies, and thus are energetically feasible. The amount of CaO present is 24.9/56.0 = 0.445 moles.

Amount of CaO required for SO_2 = 0.437 moles

Amount of CaO required for H_2SO_4 = 0.05 moles

Amount of CaO required for SiF_4 = 0.04 moles

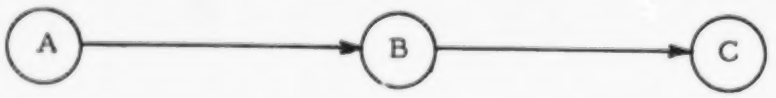
Amount of CaO required for HF, unknown

Total = 0.527 moles

Therefore, there would not be enough CaO to react with all pollutants of the acid gas form. Probably the reaction with SO_2 would take place preferably, as it has the most negative free-energy change. Another possibility is that some of the gases would combine with the emissions from the cement plant.

Exit-Gas Composition

The schematic diagram below, shows the path of the flue gases.



P = 1 atm	P_1 = -35 in. H_2O	P_2 = -50 in. H_2O
T = 336 F	T_1	T_2
V = 583,000 cfm	V_1	V_2

A = 7 Process Plants

B = Venturi Scrubber

C = Fan

To calculate T_1 and V_1, the temperature and pressure of the gas before entering the scrubber, adiabatic conditions are assumed

$$pV^{1.3} = pV^{1.3}$$

Using the relationships for adiabatic flow,

$$V_1 = 627,000 \text{ acfm}, \quad T_1 = 320 \text{ F}$$

The next step is to calculate the quantity of water picked up in the scrubber by the flue gases.

Assume water is available at 75 F, and that the specific heat of the gas approximates that of air (7 Btu/mole F).

Vapor pressure of water = 2.22 cm Hg

Latent heat of vaporization = 1050 Btu/lb.

$$P_2 = \frac{405-50}{405} = 0.875 \text{ atm}$$

Let w = weight of water evaporated into the gas stream.

A heat balance then gives:

$$1050\ w = 1010 \text{ moles of gas} \times 7 \times (780 - T_2) \ .$$

Also

$$\frac{2.22}{76} \times 2118 \times V_2 = \frac{w}{18} \times 1544 \times T_2 \ .$$

And

$$V_2 = V_{gas} + V_{water}$$

$$= 1010 \times \frac{359 \times T_2}{492} + \frac{w}{18} \times \frac{359 \times T_2}{492}$$

There are 3 equations and 3 unknowns, T_2, V_2, and w.

Solving, one obtains T_2 = 240 F

$$w = 540 \text{ lb/min (30 moles/min)}$$

$$V_2 = 530,000 \text{ cfm (394,000 scfm)}$$

The composition of the exit gases are then easily calculated.